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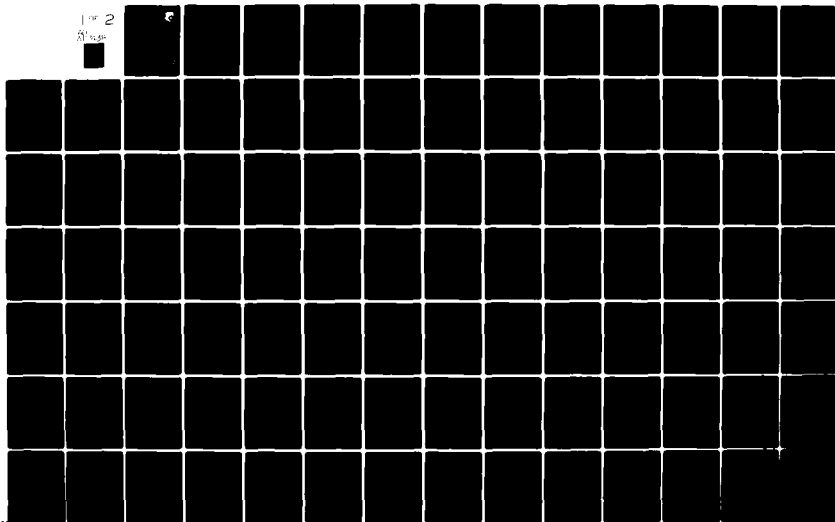
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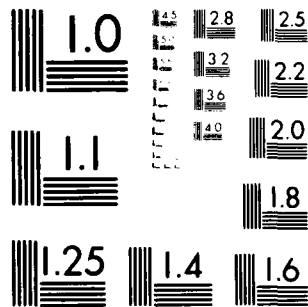
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Part V

AN EXPLORATORY RESEARCH AND DEVELOPMENT PROGRAM  
LEADING TO SPECIFICATIONS FOR AVIATION TURBINE  
FUEL FROM WHOLE CRUDE SHALE OIL



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March 1982

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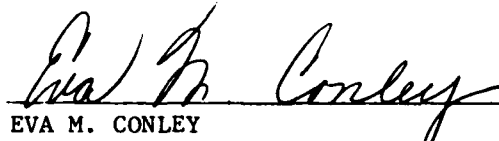
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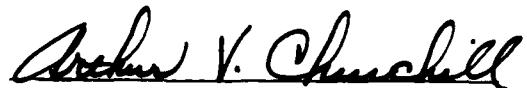
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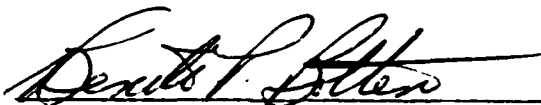
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Paraho Shale Oil	Hydrorefining	Processing Schemes
Occidental Shale Oil	Hydrogen Chloride Extraction	Investment Costs
JP-4 Jet Fuel	Basic Nitrogen	Operating Costs
JP-8 Jet Fuel	Neutral Nitrogen	Product Costs.
20. ABSTRACT (Continue on reverse side if necessary and identify by block number) A computer model of Sun Tech's upgrading concept for converting 100,000 BPSD of raw Occidental shale oil into aviation turbine fuels has been developed. Using economic guidelines provided by the U.S. Air Force, the total liquid product cost when maximizing JP-4 jet fuel was \$1.22/gallon; \$1.24/gallon when maximizing JP-8 jet fuel; and \$1.19/gallon when producing JP-4 plus other fuels. Sensitivity analysis showed that the price of raw shale oil had the greatest impact on total liquid product costs.		

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## FOREWORD

This final report presents the results of work performed by Sun Tech for the United States Air Force under contract F33615-78-C-2024.

The program is sponsored by the Aero Propulsion Laboratory, Air Force Wright Aeronautical Laboratory, Wright-Patterson AFB, Ohio, under Project 2480, Task 00 and Work Unit 01. Ms. Eva M. Conley/AFWAL/POSF, was the assigned Air Force Project Engineer.

Contract work reported herein was performed during the period of 2 January 1979 to 2 January 1981 under the direction of Dr. Abraham Schneider, Scientific Advisor, Sun Tech, Inc. and from 2 January 1981 through 1 February 1982 by Henry E. Reif. This report was released by the authors in February 1982.

Sun Tech's program managers wish to express their appreciation to Major D. D. Potter, USAF, and Lt. E. N. Coppola, USAF for their help in formulating the economic assumptions upon which the financial aspects of the program are based and to Dr. Herbert Lander and Ms. Eva M. Conley, for their assistance in overcoming administrative and logistical problems associated with this project.

The authors wish to thank E. J. Janoski for his contributions in the area of HCl extraction, J. J. vanVenrooy for pilot plant operations, G. F. Frey for assistance in estimating plant investment and operating costs, and A. Macris and J. W. Ruth for economic optimization.

This report is part V of the five planned parts covering the exploratory research and development program leading to specifications for military fuels from whole crude shale oil. Part I, "Preliminary Process Analyses" evaluated three different technically feasible processing schemes proposed by Sun Tech, Inc., for converting 100,000 BPCD of raw Paraho shale oil into military turbine fuels. Part II, "Process Variable Analyses and Laboratory Sample Production", incorporated pilot plant process data for three design bases for manufacturing military fuels from raw Occidental shale oil. Part III, "Production of 300 Barrels of JP-4 Turbine Fuel From Geokinetics Shale Oil" reports the results of the program carried out at Hydrocarbon Research, Inc., Lawrenceville, N.J., laboratory in their 30 barrel per day process demonstration unit under sub contract to Sun Tech. Part IV, "Production of Samples of Military Fuels From Raw Shale Oils" describes the production of component test samples of military fuels from both Occidental and Paraho shale oils.



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## LIST OF SYMBOLS AND ABBREVIATIONS

### SYMBOLS

Bbl/SD	Barrels per Stream Day
BTU/lb	British Thermal Units per Pound
¢/gal	Cents per Gallon
°F	Degrees Fahrenheit
\$/Bbl	Dollars per Barrel
\$/CD	Dollars per Calendar Day
\$/gal	Dollars per Gallon
\$/SDB	Dollars per Stream Day Barrel
Lbs/Bbl	Pounds per Barrel
Lbs/hr	Pounds per Hour
Lbs/SD	Pounds per Stream Day
Mol.%	Mole percent
SCF/Bbl	Standard Cubic Feet per Barrel
SCF H <sub>2</sub>	Standard Cubic Feet Hydrogen
SCF PSD	Standard Cubic Feet per Stream Day
STSD	Short Tons per Stream Day
Vol.%	Volume percent
Wt.%	Weight percent

### ABBREVIATIONS

AGO	Atmospheric Gas Oil
API	American Petroleum Institute
ASTM	American Society of Testing and Materials
Bbl	Barrel

LIST OF SYMBOLS AND ABBREVIATIONS (Continued)

ABBREVIATIONS

BPCD	Barrels per Calendar Day
BPSD	Barrels per Stream Day
BR	Boiling Range
BTMS	Bottoms
CO	Carbon Monoxide
CO <sub>2</sub>	Carbon Dioxide
cs	Centistokes
DMF	n, n-Dimethylformamide
FOE	Fuel Oil Equivalent
FCC	Fluid Catalytic Cracking
H <sub>2</sub>	Hydrogen Gas
HCl	Anhydrous Hydrogen Chloride
HP Sep	High Pressure Separator
H <sub>2</sub> S	Hydrogen Sulfide Gas
IRR	Interest Rate of Return
KV	Kinematic Viscosity
LHSV	Liquid Hourly Space Velocity
LP	Linear Program
LP Sep	Low Pressure Separator
Mo	Molybdenum
N <sub>2</sub>	Nitrogen Gas
NA	Not Available
NH <sub>3</sub>	Ammonia Gas



LIST OF SYMBOLS AND ABBREVIATIONS (Continued)

ABBREVIATIONS

NH <sub>4</sub> Cl	Ammonium Chloride
Ni	Nickel
N <sub>T</sub>	Total Nitrogen
O <sub>2</sub>	Oxygen Gas
ppm	Parts per Million by Weight
psia	Pounds per Square Inch Absolute Pressure
psig	Pounds per Square Inch Gage Pressure
R-1	First Reactor
R-2	Second Reactor
S	Sulfur
SCF	Standard Cubic Feet
SUS	Saybolt Universal Seconds
TBP	True Boiling Point Distillation
TPO	Texaco Partial Oxidation Process
VGO	Vacuum Gas Oil
VI	Viscosity Index
WWT Plant	Waste Water Treatment Plant

## SUMMARY

This report covers work performed by Sun Tech, Inc. under our contract with the United States Air Force. Phases I through III have been reported earlier in separate interim reports. In Phase IV (reported here for the first time) the objectives were to establish by computer modeling, the economically optimum processing schemes and plant capacities based on the analytical and experimental data from Phases I, II and III.

Based on the pilot plant work, Sun Tech's processing concept is viable. Using this processing concept for refining raw Occidental shale oil and the economic guidelines provided by the USAF for Phase IV, an LP computer program was developed. Due to non-linear yield effects, especially in the HCl extraction process, the optimization was performed using available experimental processing options. The results from a case-study approach were: 1) the optimal scheme for maximum JP-4 and JP-8 production was with the raw shale oil main hydrotreater operating at 2200 ppm total nitrogen content ( $N_T$ ) in the effluent; 2) the optimal scheme for the JP-4 and other fuels option was with the raw shale oil main hydrotreater operating at 6400 ppm  $N_T$  in the effluent.

Detailed process flow sheets of the major process equipment and operating conditions for the three optimal processing schemes were determined. Hydrogen consumption was 2584 SCF/Bbl of raw shale oil for maximum JP-4 production; 2363 SCF/Bbl for maximum JP-8 production; and 1960 SCF/Bbl for the JP-4 and other fuels case. Overall refinery thermal efficiency

varied from 81% for the maximum JP-8 production to 87% for the JP-4 and other fuels case. From overall economic analysis based on the Air Force guidelines we found that the price of liquid products in the maximum JP-4 case was 1.22 \$/gal, for the maximum JP-8 case was 1.24 \$/gal and for the JP-4 and other fuels case was 1.19 \$/gal with raw shale oil priced at \$40 per barrel (0.95\$/gal.). Sensitivity analyses on the economic variables, using a computer program, showed that the price of raw shale oil had the largest impact on product prices, that changes in the discounted cash flow rate and variation in capital expenditure and finally annual interest rate for working capital had only a small impact on fuel prices. Examining a more realistic scenario, where both working and plant investment capital has to be borrowed at an annual rate of 15%, we found that the product prices increase by about 9.5 cents per gallon.

## SECTION I

### INTRODUCTION

#### 1. Background

In previous work sponsored by the Department of Energy and the Department of Defense, Robinson demonstrated that specification quality JP-5 could be produced from raw Paraho shale oil.<sup>(1)</sup> The manufacturing processing sequence consisted of the following three steps:

- (1) Hydrotreating raw shale oil to lower its non-hydrocarbon content and to increase the hydrogen to carbon ratio,
- (2) Fractionating the hydrotreated shale oil into the desired boiling range fractions, and finally
- (3) Acid and clay treating to meet thermal and storage stability requirements.

A variation of this processing sequence was evaluated in Sun Tech's Phase I Base Case. The variations consisted of the following processing steps:

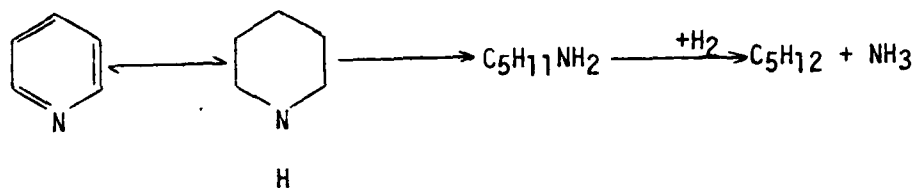
- (1) Increasing hydrotreating severity to lower the total nitrogen content of the reactor effluent to 300 ppm vs. 3000 ppm in the reference,
- (2) Washing of the hydrotreated shale oil with 80% sulfuric acid to provide product stability and,
- (3) Final distillation into the desired product boiling ranges.

Sun Tech's alternate processing concepts are based on in-house experience. Initially raw shale oil is hydrotreated, as in the Base Case, but at lower severity, then followed by distillation. The heavy distillate fraction is extracted to further reduce its nitrogen content. The nitrogen content of the raffinate phase is now reduced sufficiently for

charging directly into a hydrocracker. The extract phase, which is rich in heteroatoms, is then used to manufacture hydrogen by partial oxidation. Through the use of moderate hydrogenation severity, hydrogen is conserved.

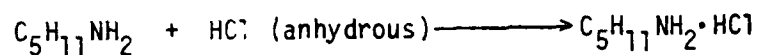
Whole crude shale oil typically contains approximately 2 weight percent of nitrogen of which 50 to 70 weight percent is in the basic form. The major portion of the nitrogen is present in five and six member rings which are unsaturated and polycyclic in nature. Before crude shale oil can be processed into transportation fuels using conventional petroleum catalytic conversion processes, the nitrogen level must be significantly reduced or essentially eliminated to avoid poisoning the acid function of catalysts. Removal of this nitrogen can be accomplished by hydrodenitrogenation as described by Cocchetto and Satterfield.<sup>(2)</sup> Nitrogen, for the most part, is present as heterocyclic compounds. It is reduced to ammonia and removed as such or the heterocyclic compounds are saturated to basic nitrogen structures. All compounds are then extracted with a mineral acid, such as anhydrous HCl, to form an amine hydrochloride. Most of the amine hydrochlorides are insoluble in hydrocarbons and form a dense and viscous liquid phase which separates from the hydrocarbons in the system.

It has been reported by Dinneen<sup>(3)</sup> that fractions of Colorado shale oil contain pyridines, indoles, quinolines, tetrahydroquinolines and more complex structures. Hydrodenitrogenation of these compounds as described by McIlvried et al., generally proceeds by first saturating the nitrogen bearing ring, breaking the carbon-nitrogen bond and then removing the nitrogen from the amine as ammonia.<sup>(4)</sup>



It can be seen from the above model equation that hydrotreating converts the unsaturated heterocyclic compound (pyridene) to the saturated structure (piperidine) or the aliphatic amylamine.

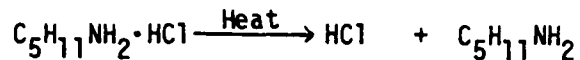
The addition of anhydrous HCl can form the hydrochloride salt of either one or both of the nitrogen containing compounds.



It can be seen that hydrogen would be conserved by not proceeding all the way to form ammonia.

Examination of the amine hydrochloride extract showed the presence of both basic and neutral nitrogen. The ratio of basic nitrogen to total nitrogen was 0.775. The ratio suggests a bonus of an additional 30% removal of nitrogen per chlorine atom indicating that some molecules contain both basic and neutral nitrogen.

Decomposition of the extract releases HCl and the recovered extract can be used for manufacturing hydrogen by partial oxidation. This process can be represented by the following equations:



Downstream processing converts the CO to H<sub>2</sub> and CO<sub>2</sub> via the water-gas shift reaction.

## 2. Overall Objectives

The overall objectives of Sun Tech's Phase IV economic optimization studies were to:

- (1) Establish by computer modeling the economically optimum processing scheme and plant capacities based on analytical and experimental data from Phases I, II, and III.
- (2) Determine the economic and yield trade-offs for producing JP-4 or JP-8 turbine fuels as primary products.
- (3) Provide detailed process flow sheets of the major process equipment and operating conditions for the optimized shale oil processing scheme.
- (4) Estimate external resources required for each process investigated--i.e., water, electricity, and hydrogen.
- (5) Define remaining problems and/or uncertainties.

## SECTION II

### SUN TECH'S UPGRADING CONCEPT

Sun Tech's processing concept for refining raw Occidental shale oil consists of six distinct steps: (1) hydrotreating the whole shale oil to partially reduce the high total nitrogen content (and convert some neutral nitrogen to basic nitrogen), while minimizing hydrogen consumption; (2) distilling the hydrotreated product into appropriate fractions for additional processing; (3) rehydrotreating the light distillate fraction to meet product specifications; (4) treating the wide boiling distillate fraction with anhydrous hydrogen chloride which yields a raffinate and extract phase--the nitrogen content in the HCl raffinate is lowered and concentrated in the extract phase; (5) thermally decomposing the HCl extract to recover anhydrous hydrogen chloride--the recovered HCl-free nitrogen-rich extract fraction is used for generating hydrogen by partial oxidation; and (6) hydrocracking the raffinate fraction to maximize the yield of aviation turbine fuels. In Phase I, "Preliminary Process Analyses" three different technically feasible processing schemes proposed by Sun Tech, Inc., for converting 100,000 BPCD of raw Paraho shale oil into military turbine fuels were evaluated. Phase II, "Process Variable Analyses and Laboratory Sample Production", incorporated pilot plant process data for three design bases for manufacturing military fuels from raw Occidental shale oil. In Phase III, total of 475 gallons specification aviation turbine fuels were prepared from Occidental shale oil--170 gallons of JP-4, 150 gallons of JP-5, and 155 gallons of JP-8. A block flow diagram of Sun Tech's upgrading process is shown in Figure 1.



A modification of Sun Tech's processing scheme was employed in processing Paraho shale oil. The modified processing route involves severely hydro-treating the raw shale oil followed by hydrocracking the gas oil fraction. Five 5-gallon samples of specification military fuels were produced from Paraho shale oil--JP-4, JP-5, and JP-8 jet fuels, along with Diesel Fuel No. 2 and Diesel Fuel Marine. A block flow diagram for preparing military fuels from Paraho shale oil is given in Figure 2.

### 1. Shale Oil Characterizations

Sun Tech has evaluated two different shale oils during the course of its work with the United States Air Force. The predominant feedstock used was Occidental modified in-situ shale oil. Paraho shale oil obtained from a directly heated surface retort was also evaluated. Table 1 presents inspections and analyses for both Occidental and Paraho shale oils. Occidental can be processed using less severe conditions than required for Paraho based on boiling range, nitrogen, sulfur, and hydrogen contents. Both shale oils contain significant quantities of arsenic not found in conventional petroleum and the nitrogen and oxygen contents of raw shale oil are also higher than those found in conventional petroleum.

### 2. Processing Description and Configuration

#### a. Feedstock Preparation

The raw shale oil is heated to 175°F and is allowed to stand. A water layer is separated out and most of the fines are removed. Finally, the

dewatered shale oil is pumped through a 5 micron Cuno filter before charging to the raw shale oil hydrotreater.

b. Raw Shale Oil Hydrotreater

A simplified flow diagram of the raw shale oil hydrotreater is shown in Figure 3. Dewatered and desilted shale oil, stream 1, is pumped to reactor pressure and split into two parallel streams to be fed to guard reactors, R-100 A & B. The shale oil to each guard reactor is combined with make-up and recycle hydrogen, streams 3 and 4. The mixed feed is heated to guard reactor inlet temperature in the feed/effluent heat exchanger E-100. Guard reactor effluent is heated to hydrotreater reactor inlet temperature by fired heater H-100 and quench gas is injected between catalyst beds to control temperature rise.

Hydrotreater reactor effluent is cooled by exchange with reactor feed and air cooler E-101 to 275°F. The mixed phase is flashed in V-101 and separated into hydrocarbon vapor and liquid phases. After further cooling and separation in E-102 and V-102, wash water is combined with the vapor phase to remove ammonia and some hydrogen sulfide.

The cold effluent is separated into a hydrogen rich gas stream, a sour water stream and a hydrocarbon liquid stream in high pressure separator V-104. The gas stream, processed in T-100 for  $\text{NH}_3$  and  $\text{H}_2\text{S}$  removal, is recycled to the reactors. The sour water is sent to waste water treatment and the hydrocarbon liquid from V-101 is combined with liquid from V-102 and sent to low pressure separator V-103.

Hydrocarbon is flashed at 150 psig in V-103 and separated into vapor and liquid phases. The vapor is cooled in E-103. The cold effluent is separated into a vapor stream and a hydrocarbon liquid stream in V-104. The vapor phase is sent to fuel gas and the hydrocarbon streams from V-103 and V-104 are combined as product and sent to fractionation.

#### c. Hydrotreated Shale Oil Fractionation

A simplified flow diagram of the hydrotreated shale oil atmospheric and vacuum distillation units is given in Figure 4. Hydrotreated shale oil, stream 2, is heated by fired heater H-101 before being fed to the atmospheric fractionator T-101. The bottoms from the atmospheric column are heated by fired heater H-102 before entering the vacuum tower, T-104. The following streams from the fractionation plant are obtained:

<u>Stream No.</u>	<u>Description</u>
7	Light Ends to H <sub>2</sub> Plant
8	C <sub>4</sub> -180°F for JP-4 Jet Fuel Blending
9	Naphtha to Naphtha Hydrotreater
10	Atmospheric and Vacuum Gas Oils to HCl Extraction
11	1000°F+ Bottoms to TPO, fuel or fuel blending

The fractionation cut points can be varied depending on the type of operation, JP-4 or JP-8 production.

#### d. Naphtha Hydrotreater

The purpose of the naphtha hydrotreater is to cleanup the light distillate from the atmospheric distillation column in order to meet final

product specifications. A simplified flow diagram of the naphtha hydro-treater and fractionator is given in Figure 5. Naphtha feedstock, stream 9, is combined with makeup and recycle hydrogen. The mixed feed is heated to reactor inlet temperature in feed/effluent heat exchanger E-107 and fired heater H-103. Hydrogen quench gas is injected between catalyst beds in reactor R-102 to control temperature rise. Hydrocarbon is flashed at 150 psig in V-109 and separated into vapor and liquid phases. The vapor is washed with water for  $\text{NH}_3$  and  $\text{H}_2\text{S}$  removal, before being sent to the recycle compressor, C-101. The hydrocarbon streams from V-110 and V-111 are combined as liquid product and sent to the depropanizer, T-106. If JP-4 is being produced, fractionator T-107 is not required. In JP-8 production, the hydrotreated products consist of a  $\text{C}_4$ -290°F gasoline blendstock and the JP-8 product, that is the 290-550°F boiling range fraction.

#### e. HCl Extraction

The purpose of Sun Tech's HCl extraction step is to remove much of the remaining nitrogen compounds found in the atmospheric and vacuum gas oils obtained from hydrotreated shale oil. Through the use of this step, less hydrogen is needed in the overall refinery.

A simplified flow diagram of Sun Tech's continuous anhydrous HCl extraction plant is shown in Figure 6. Vacuum dried gas oil, stream 10, enters the top of the HCl absorption column, T-108, where it is contacted countercurrently with makeup and recycle HCl. The reduced nitrogen raffinate is separated from the extract in the adduct settler, V-115. The raffinate is water washed in column T-109 before being sent to hydrocracking as

stream 20. Recycle HCl is recovered from the extract phase in the HCl flash drum, V-114, and combined with makeup HCl for use in the HCl absorption column. The thermally decomposed extract phase, which is rich in heteroatoms, is used to manufacture hydrogen by partial oxidation.

#### f. Raffinate Hydrocracking

Raffinate hydrocracking is employed in Sun Tech's shale oil upgrading process to increase the yield of aviation turbine fuels. Figure 7 presents a simplified flow diagram of the raffinate hydrocracker. Hydrocracking the raffinate from the HCl extraction step required two reactors -- R-103 to partially saturate the aromatics and to remove the remaining nitrogen and sulfur compounds from the raffinate, and R-104 for molecular weight reduction to produce aviation turbine fuels.

Raffinate feedstock, stream 20, is heated in feed/effluent exchangers E-116 and E-114 and combined with makeup and recycle hydrogen. The mixed feed is sent to reactor R-103 for nitrogen and sulfur removal. Hydrogen quench gas is injected between catalyst beds to limit temperature rise. Water is injected in the effluent from reactor R-103 to remove ammonium chloride, ammonia, and hydrogen sulfide. The water washed effluent is combined with hydrogen and sent to the hydrocracking reactor, R-104.

A simplified flow diagram of the hydrocracker fractionation plant is given in Figure 8. Fractionator cut points depend on the product slate desired. A recycle drag stream, may be required due to the buildup of aromatics or wax in the recycle oil, stream 21. In the JP-4 plus other

fuels case, there is no recycle oil sent back to the hydrocracking reactor R-104. Additional products include a 490-675°F boiling Diesel Fuel #2 blendstock, and 675°F+ bottoms for heavy fuel blending.

#### g. Hydrogen Manufacturing Processes

##### 1) General

Two different hydrogen manufacturing units are incorporated in the proposed processing schemes to utilize the available feedstocks. One unit operates on a light hydrocarbon feed,  $C_1$ - $C_4$  co-products from hydrotreating and hydrocracking steps. The second unit produces hydrogen by partial oxidation of heavy hydrocarbon feeds, i.e. decomposed HCl extract, 1000°F+ hydrotreated bottoms or raw shale oil.

The manufacture of hydrogen by steam reforming of light hydrocarbons is a well established process and will not be discussed further since yields, hydrogen purity and operating costs are well known. We have assumed that raw shale oil can be used as a fuel to steam reforming furnaces. Since this procedure has not been practiced commercially, the validity of this assumption is not certain.

##### 2) Manufacture of Hydrogen by Partial Oxidation

Manufacture of hydrogen from heavy feeds containing high concentrations of heteroatoms, such as sulfur, nitrogen, oxygen and chlorine required assistance from the process licensor to insure that the process would be operable with the feeds proposed. In addition, information was needed for estimating yields and operating costs. The Texaco Partial Oxidation process (TPO) was selected for our application. Based on the analysis of

our poorest quality feedstock, Texaco Development Corporation confirmed that their process could operate on decomposed HCl extract. They also furnished estimates of feed and utility requirements plus product gas composition. From the information supplied by Texaco<sup>(5)</sup> and the literature<sup>(6)</sup> we estimated plant investments and operating costs for each feedstock. These data were used in the refinery L.P. for maximizing various product yields and balancing hydrogen manufacture.

In the Texaco Partial Oxidation process gases generated in the partial oxidation reactor consist mainly of carbon monoxide and hydrogen. The exit gases are first scrubbed with naphtha to extract carbon particles (for recycling to the oxidation reactor) before entering the downstream conversion and purification system. Here, carbon monoxide is converted to hydrogen and carbon dioxide by a catalytic water gas shift reaction and the carbon dioxide is removed by extraction with methanol. Finally any residual carbon oxides remaining in the treated gas stream are catalytically converted to methane. Ultimate hydrogen purity ranges from 97 to 99 mol %.

From the process information furnished by Texaco along with that from the literature a set of guidelines was developed for estimating hydrogen yields and purities via TPO. Hydrogen produced from the proposed feedstocks were calculated using these guidelines and gave the following results:

- 1) 1000°F+ bottoms from hydrotreated Occidental shale oil yields the most hydrogen per barrel - 15,800 SCF/H<sub>2</sub> @ 98.2 mole % purity.

- 2) Decomposed nitrogen extracts from either hydrogenated Paraho or Occidental shale oils are essentially equal, but yield about 400 SCF less hydrogen per barrel of feed than the 1000°F+ bottoms-about 15,400 SCF/H<sub>2</sub> @ 97.7 mole % purity.
- 3) Raw shale oils (Paraho, Occidental or Geokinetics) are essentially equal, but they yield about 700 SCF less hydrogen per barrel than the nitrogen extracts - about 14,700 SCF/H<sub>2</sub> @ 97.9 mole % purity.

Any of the above feedstocks would be suitable for use in the TPO process. These estimated hydrogen yields were used in the refinery math model, for providing a basis for selecting and ranking feedstocks going to the hydrogen plant and for optimizing the depth of hydrogenation in the raw shale oil hydrotreater.

Texaco Development Corporation's data indicates that the major portion of nitrogen in the feedstock appears as elemental nitrogen in the exit gases and the remainder is converted to ammonia. Combined chlorine remaining in the extract feedstock is converted to hydrogen chloride which reacts with ammonia in the exit gases. The resulting ammonium chloride is extracted by the water scrubber.

The depth of hydrogenation in the raw shale oil hydrotreater controls the yield of nitrogen extract and 1000°F+ bottoms which in turn are used to manufacture hydrogen by TPO. Pilot plant data obtained at three different hydrotreating depths were used to estimate extract yields, physical properties and elemental analysis of decomposed extracts. Elemental



analyses showed only minor changes in the extract compositions from a particular shale oil hydrogenated to varying depths.

Hydrogen yield and purity data for the various feedstocks are summarized in Table 2. Steam and oxygen requirements are also shown. The operating pressure selected for the TPO plant was 950 psig. Estimated utility requirements are given in Table 3.

#### h. Waste Water Treating Process

The Chevron Waste Water Treating (WWT) Process is a patented process for treating foul water streams from petroleum refineries and synthetic fuel plants to: a) recover and separate high purity ammonia and hydrogen sulfide; and b) to recover clean water suitable for reuse or for discharge. Investment and operating costs for the WWT plant have been provided to Sun Tech by Chevron Research Company.

### SECTION III

#### PROCESS VARIABLE ANALYSIS

During Phase II and Phase III of our contract with the United States Air Force, Sun Tech's shale oil upgrading concept was evaluated in the laboratory and pilot plant. Detailed description of this work can be found in the interim reports, "Part II - Process Variable Analysis and Laboratory Sample Production"<sup>(7)</sup> and "Part IV - Production of Samples of Military Fuels from Raw Shale Oils".<sup>(8)</sup>

##### 1. Raw Shale Oil Hydrotreater and Distillation Units

A simplified flow diagram of the raw shale oil hydrotreater and distillation units is shown in Figure 9. The use of guard reactors is necessary to remove arsenic and iron, as well as to saturate olefins in the feed. A vacuum still is used to produce a gas oil fraction with a 1000°F end point. The waxy nature of the 1000°F+ bottoms precludes its use in the HCl extraction step due to the formation of emulsions. Operating conditions used in the raw shale oil hydrotreater to yield a liquid product containing 5000 ppm total nitrogen are given in Table 4.

A total nitrogen content of 5000 ppm in the hydrotreated product was chosen to produce sufficient extract for hydrogen manufacture by partial oxidation. Two additional levels of hydrogenation severity at 2200 and 6400 ppm total nitrogen content in the reactor effluent were also evaluated. All three levels have been incorporated in Sun Tech's math model for process optimization. Operating conditions required to obtain these

additional levels of nitrogen in the reactor effluent are also given in Table 4.

Material balance summaries for the main hydrotreater and distillation units are given in Tables 5, 7, and 9 for each of the reactor effluent nitrogen levels. Significant quantities of ammonia, water, and hydrogen sulfide are produced during hydrogenation. Cut points for the distillation unit are varied depending on the type of operation, JP-4 production or JP-8 production. Product inspections on the streams from the main hydrotreater distillation units are shown in Tables 6, 8, and 10. Nitrogen, sulfur, and aromatics contents increase with increasing boiling range. Very little material is found boiling below 250°F in the hydro-treated product.

## 2. Naphtha Hydrotreater

The purpose of the naphtha hydrotreater, shown schematically in Figure 10, is to clean up the light distillate from the atmospheric distillation unit to meet product specifications. The effluent is passed through a product stripper (not shown) before blending into final products. Operating conditions used in the naphtha hydrotreater are given in Tables 11, 14, and 17. Material balance summaries for the JP-4 operations are presented in Tables 12, 15, and 18. Material balance summaries for the JP-8 operations are given in Tables 13, 16, and 19. In the JP-4 case, feedstock and product boiling ranges are 180-490°F. In the JP-8 case, the feedstock boiling range is 180-550°F; however, the hydrotreated products

consisted of a C<sub>4</sub>-290°F gasoline blendstock and the 290-550°F JP-8 fraction.

### 3. Extraction Processes

Three alternate processes for removal of nitrogen compounds remaining in mildly hydrotreated shale oil were evaluated. DMF and methanol appear to be about equal for extracting nitrogen compounds from light distillates (700°F end point) derived from mildly hydrotreated Occidental shale oil. These solvents would be useful for removing nitrogen compounds in the JP-4 through #2 diesel fuel (DF-2) boiling range. Above 700°F, these solvents were only marginally effective exhibiting poor selectivity for nitrogen removal. HCl extraction of the 450-1000°F distillate fractions of hydrotreated shale oil was more effective for removal of nitrogen containing compounds than either DMF or methanol extraction. Therefore, HCl extraction was the process chosen to remove nitrogen compounds from high boiling fractions of mildly hydrotreated Occidental shale oil.

#### a. HCl Extraction

Pilot plant HCl treating was carried out batchwise. Due to the smooth operation of these runs, we feel that the process can be readily adapted to continuous operation and achieve similar results. A schematic flow diagram of a continuous HCl extraction plant is shown in Figure 11.

Operating conditions for HCl treating and material balance summaries for the JP-4 operation are presented in Tables 20 through 23. Here the gas oil feedstock has a 490-1000°F boiling range and a total nitrogen content

range varying from 2400 to 6887 ppm. Tables 24 through 27 give the operating conditions and material balance summaries for the JP-8 operation. In this case, the gas oil feedstock has a 550-1000°F boiling range and a total nitrogen content varying from 2400 to 7100 ppm. Considerable amounts of chlorides remain in both the raffinate and decomposed adduct. There is a 0.1 volume % loss of raffinate and a 5 weight % loss of anhydrous HCl in the water washing step.

#### 4. Raffinate Hydrocracking

A single stage hydrocracker is shown in Figure 12. Reactor R-1 is used to clean up the raffinate feed before it enters the main hydrocracking reactor R-2 where most of the hydrocracking takes place. The fractions taken off the distillation tower can be varied. Extinction recycle of the distillation bottoms is optional.

Table 28, 30 and 32 present the hydrocracker operating conditions for maximum production of JP-4 jet fuel. Originally, we intended to use a proprietary hydrocracking catalyst with which we have had experience. We were barred from using this catalyst for shale oil applications. After screening three non-proprietary catalysts, a nickel tungsten catalyst designated "B" was selected for this operation. Material balance summaries for the maximum JP-4 operation are given in Table 29, 31, and 33. Ammonium chloride formed during the R-1 hydrotreating reaction is removed by the injection of water before the high pressure separator. High yields of JP-4 jet fuel are obtained.

Hydrocracker operating conditions and a material balance summaries for maximizing JP-8 production are presented in Tables 34 through 39. In this case, a portion of the total liquid product is C<sub>4</sub>-290°F gasoline blendstock. The remainder of the liquid product is JP-8 jet fuel having a 290-550°F boiling range. Here we have the option of recycling the 550°F+ bottoms to extinction or taking a drag stream. Again, ammonium chloride is removed by injecting water after the R-1 hydrotreater.

Tables 40 through 45 summarize the Phase II hydrocracker operation for production of JP-4 and other fuels. In this operation there is no recycle stream to the R-2 hydrocracker (once-through operation). In addition to JP-4 jet fuel, #2 diesel fuel (DF-2), and a 675°F+ bottoms fuel oil are produced. Since there is no recycle oil to the R-2 hydrocracker, chemical consumption of hydrogen is significantly lower than in the maximum JP-4 case.

#### 5. Catalyst Life Studies

Considerable effort was expended in selecting and evaluating non-proprietary catalysts for use in various catalytic processing units. In order to proceed with the overall economic optimization work, catalyst life estimates were developed for the R-1 guard reactor and the R-2 hydrotreater reactor in the raw shale oil hydrotreater based on pilot plant catalyst aging runs. Catalyst life estimates were also estimated for the naphtha hydrotreater and the gas oil hydrocracker.

a. Raw Shale Oil Hydrotreater

A two reactor isothermal pilot plant was employed to determine catalyst aging characteristics in the R-1 guard reactor and the R-2 hydrotreater reactor. The catalyst aging curve, Figure 13, shows that after the loss of the initial high activity characteristic of fresh catalysts, the temperature required in the R-2 catalyst bed to hydrotreat whole Occidental shale oil to 5000 ppm total nitrogen in the reactor effluent remained essentially constant. Almost four months of successful life-testing was accumulated with Occidental shale oil. Catalyst activity tests were run periodically to determine the average catalyst temperature required to produce 5000 ppm total nitrogen in the reactor effluent. Most of the on-stream time employed more severe operating conditions producing 2200 ppm total nitrogen. A minor portion of the time produced material containing 6400 ppm total nitrogen. The R-1 guard reactor catalyst bed was kept at a maximum temperature of 650°F during the seven month catalyst life test.

Using the same catalyst loading that had accumulated almost four months of life with Occidental shale oil, an additional two month life test with Paraho shale oil was completed. Since the Paraho feed contained 2.13 wt.% total nitrogen as opposed to the 1.46 wt.% total nitrogen content found in Occidental shale oil, a 50°F increase in R-2 average catalyst bed temperature was required to yield a hydrotreated product containing 5000 ppm total nitrogen (see Figure 14). At this point the feed was changed back to Occidental shale oil and the activity checked. During the two months the unit was operated on Paraho shale oil, the catalyst activity aged 10°F. Based on the stable aging characteristics of the

catalyst in R-2, a life expectancy of 1 year is projected; for R-1 we project a 6-month catalyst life. Arsenic content in the R-1 effluent varied between 0 and 1 ppm. Finally, an additional one-month long run with Occidental shale oil was made employing severe operating conditions producing less than 5 ppm total nitrogen in the reactor effluent. During this period of severe operation, some catalyst activity loss was apparent.

#### b. Naphtha Hydrotreater

Based on feedstock composition, unit operating conditions, and Sun Tech's experience with commercial petroleum units, a 2.5 year catalyst life is estimated for the naphtha hydrotreater, when treating naphtha from the 2200 ppm nitrogen syncrude; a 2.0 year catalyst life when treating naphtha from the 5000 ppm nitrogen syncrude; and a 1.5 year catalyst life when treating naphtha from the 6400 ppm nitrogen syncrude.

#### c. Raffinate Hydrocracker

Using the same criteria described above, the estimated catalyst life for the R-1 hydrotreater was 6 months. The estimated catalyst life for the R-2 hydrocracker was 1.25 years when maximizing JP-4 jet fuel and 1.8 years when maximizing JP-8 jet fuel or when producing JP-4 plus other fuels.

### 6. Product Inspections

Specification quality JP-4, JP-8, DF-2, and  $C_4$ -290°F gasoline blend-stock can be produced by Sun Tech's process to upgrade raw Occidental shale oil. Product inspections are presented in Table 46. Essentially



complete removal of nitrogen and sulfur is obtained. The blended heavy fuel consists of the 1000°F+ bottoms from the vacuum distillation unit blended with the 675°F+ fuel oil derived from the JP-4 plus other fuels operation. Some nitrogen and sulfur remain in the blended heavy fuel.

Section IV  
REFINERY OPTIMIZATION

1. Purpose

The purpose of Phase IV of the program was to develop a computer model of Sun Tech's shale oil upgrading process for use in optimizing the process to maximize the production of either JP-4, JP-8 or JP-4 aviation turbine fuel plus other military fuels. This study utilized the analytical and experimental data generated in Phases II and III of the program.

2. Refinery Design Basis

The refinery configuration used for this optimization study consisted of the following major process units:

- 1) Raw shale oil hydrotreater and hydrogen sulfide recovery
- 2) Atmospheric and vacuum distillation
- 3) Light distillate hydrotreater
- 4) Heavy distillate hydrotreater
- 5) Hydrocracker and atmospheric distillation
- 6) Hydrogen manufactured via
  - a) Steam reforming (light hydrocarbons)
  - b) Partial oxidation (heavy feedstocks)
- 7) Waste water treating and ammonia and hydrogen sulfide recovery
- 8) Sulfur recovery

The refinery was designed to process 100,000 BPSD of raw Occidental shale oil. After the raw shale oil is dewatered and desilted, it is hydro-treated and fractionated. Capacities of the units downstream of the fractionator vary slightly due to the changes in severity of the processing step required to optimize a specific product or slate of products. Only three severities in the raw shale oil hydrotreater were studied -- 2200, 5000 and 6400 ppm total nitrogen levels in the liquid effluent. Since the yield of HCl extract is not linear, these three levels were evaluated to optimize the yield of each fuel.

### 3. Computer Modelling

The logic of Sun Tech's LP model is shown schematically in Figure 15. 100,000 BPSD of raw shale oil are upgraded in the sequence of processing units shown. All plants, except the hydrogen manufacturing plants, are of set size for the specific product slate option evaluated. The steam reformer processes all the light ends available, and has the option of using  $C_4$ 's as feed. The TPO plant can assume any size to close the hydrogen balance. The LP model arrives at the economically optimal size of the hydrogen manufacturing plants according to the feeds available and the operating and capital costs involved.

The first processing step is a moderate hydrotreat to reduce the high nitrogen content of raw shale oil. The unit is modeled at the three levels of effluent nitrogen contents that were studied experimentally. The liquid product is distilled to yield four cuts. The  $C_4$  - 490°F cut goes to the naphtha hydrotreater. The 490°-550°F cut can go either to

the naphtha hydrotreater or the HCl extraction plant, depending on which jet fuel product is being maximized. The 550°-1000°F gas oil cut is sent to the HCl extraction plant; and the 1000°F+ bottoms can be used for H<sub>2</sub> production in the Texaco Partial Oxidation (TPO) plant, used for refinery fuel, or can be blended into heavy fuel. Light ends and waste water containing H<sub>2</sub>S and NH<sub>3</sub> go to the steam reforming unit and the Chevron Waste Water Treating (WWT) plant, respectively. All H<sub>2</sub>S recovered is sent to the sulfur plant for conversion to elemental sulfur.

The naphtha fraction is rehydrotreated to meet product specifications. The liquid products are either sent for aviation turbine fuel blending or go directly as final products. Light ends go to the steam reformer for hydrogen manufacture. Hydrogen sulfide and ammonia are recovered from the waste water.

The gas oil fraction is treated with anhydrous HCl which yields a raffinate phase much lower in nitrogen content than the feed and a nitrogen rich extract phase. The HCl raffinate goes to the hydrocracker, while the HCl extract, after thermal decomposition to recover HCl, can be used in the TPO plant, as refinery fuel, or can be sold as final product.

The hydrocracker operates on a recycle mode to maximize the yield of JP-4 or JP-8, and on an once-through basis to yield a variety of final products, such as #2 diesel fuel, and diesel fuel marine. Liquid products are fractionated and either collected for blending or sold directly. Light ends and waste water containing H<sub>2</sub>S, NH<sub>3</sub> and NH<sub>4</sub>Cl are generated in this unit.

Light ends from all hydroprocessing units are assumed to be similar in composition and are sent to the stream reforming plant to manufacture hydrogen. The  $H_2$  balance is closed by sending raw shale oil to the TPO plant to supplement the 1000°F+ bottoms and HCl extract feedstocks. The TPO plant is modeled to use all HCl extract first, then 1000°F+ bottoms, and finally raw shale oil.

The waste water streams containing  $NH_3$  and  $H_2S$  are collected and sent to the WWT plant, where the coproducts are separated and recovered. Fuel and three grades of steam are provided to the operating units, through a boiler house, not shown in Figure 15. Raw shale oil, 1000°F+ bottoms and/or HCl extract can also be used to provide process fuel and generate steam.

Finally, the appropriate refinery streams are collected for blending into aviation turbine fuels. The final product, JP-4 or JP-8, is blended from the collected streams to meet product specifications.

In order to completely describe each unit in the shale oil refinery LP model, we require feed and yield data, operating cost and utility requirement data for each operating mode of the various units. The yield data used for our LP model were obtained during the Phase III work. Experimental data were used whenever possible. However, yield data for the hydrocracking plant were developed using Sun Tech's proprietary Hydrocracking Kinetic Math Model, which was "calibrated" for shale oil hydrocracking from the Phase II pilot plant data. The steam reformer data were developed using Sun Tech's proprietary Hydrogen Plant Math

Model. WWT and TPO yield data were obtained from the process licensors, Chevron and Texaco. All the experimental data used were obtained from the pilot plant operations using Occidental raw shale oil.

Operating costs included catalyst replacement, electricity, cooling water, chemicals and royalties for all plants are based on the best data available. Capital costs for the units were estimated by Sun Tech's Engineering Department. Some units were essentially the same for each operating strategy, and were not included in the calculation of capital cost. The capital cost for these units was included externally at the end of each cycle of evaluations.

Utility requirement data, such as fuel and steam, used in the shale oil refinery LP were those reported in the Phase II report.<sup>(7)</sup> The steam reformer fuel requirements were changed to match the predictions from the Sun Tech hydrogen plant model.

To complete our LP model we used the USAF economic guidelines shown in Table 47 to provide information on feed availability and product prices. The refinery throughput was set to be 100,000 BPSD with an additional 20,000 BPSD maximum for H<sub>2</sub> manufacture or refinery fuel.

Once the LP was developed, it allowed various alternative processing schemes to be evaluated quickly and efficiently. Using a case study approach we found that in order to maximize JP-4 and JP-8 production the raw shale oil hydrotreater had to be operated at 2200 ppm N<sub>T</sub> content in the effluent, while for the JP-4 plus other fuels case, raw shale oil

hydrotreating severity was most economical when operated at 6400 ppm. A material balance summary of the three optimized cases is presented in Table 48. The hydrogen consumption for each unit is summarized in Table 49.

The results in the optimization study were very much a function of the economic guidelines used. Pricing all the products at \$50 per barrel does not take into account the fact that some products (like JP-4) are more desirable than others (like heavy fuel). When aviation turbine fuels are maximized the highest severity of hydrotreating was found optimal, although it is the most hydrogen consuming scheme. Going to even a more severe hydrotreating option (for example 700 ppm  $N_T$  in the effluent), and thereby eliminating the HCl extraction plant did not prove to be more economical. Therefore it appears that there exists some optimal hydrotreating operation which lies between the two options we studied (the 700 and 2200 ppm  $N_T$  in the raw shale oil hydrotreater effluent).

For JP-4 and other fuels production the 6400 ppm  $N_T$  in the effluent case was optimal. This result might have been different if more realistic product pricing was used.

The 5000 ppm  $N_T$  in the raw shale oil hydrotreater effluent cases for all product slate options considered was the worst. This result was due to the production of large amounts of nitrogen rich HCl extract phase. As is seen in Figure 16, the HCl extract yield approaches the maximum for this case and therefore a minimum of HCl raffinate is produced, which in turn results in lower volumes of final products.

Since the TPO unit is one of the most expensive units in the shale oil refinery to build, we examined the possibility of replacing the TPO plant with a naphtha reforming unit. Data for the naphtha reforming unit were developed using Sun Tech's hydrogen plant model and literature data. Excess butanes and naphtha streams were available to satisfy the  $H_2$  balance for all JP-8 cases and the 6400 ppm  $N_T$  for the JP-4 case. In all circumstances though, eliminating the TPO plant left us with decomposed HCl extract as product which is of marginal quality and not desirable for use as a fuel. Also, the large utility requirements of the naphtha reformer plant in comparison to the TPO utility requirements made the inclusion of a TPO plant favorable.



## SECTION V

### ENGINEERING DESIGN BASES

Using Sun Tech's linear program, we found that the optimal processing scheme for maximum JP-4 and JP-8 production was achieved with the raw shale oil hydrotreater operating at 2200 ppm total nitrogen ( $N_T$ ) content in the effluent, while for JP-4 and other fuels production was achieved with the unit operating at 6400 ppm  $N_T$  in the effluent.

The economic guidelines used in developing the LP model and all the subsequent Phase IV economics were described in Table 47. A first quarter of 1981 cost base was used with 100% equity financing. Crude shale oil was valued at \$40/bbl and all liquid product fuels were equally valued at \$50/bbl for working capital calculations and at actual cost of the overall refinery economic studies.

Material balances around the refinery and overall thermal efficiencies for the three optimized cases were summarized in Table 48. Sun Tech's Engineering Department used the optimized downstream plant capacities to come up with capital costs for all the refinery units. The optimized plant capacities and estimated first quarter 1981 investments for the three optimized cases are summarized in Table 50. The main hydrotreater consists of two parallel units with the effluents fed to a single atmospheric and vacuum distillation plants. The gas oil hydrocracker consists of two parallel trains with their effluents combined with the effluent from the distillate hydrotreater and distilled in the same fractionator unit. The main hydrotreater is the most expensive unit accounting for

about 1/3 of the total on-site costs. The TPO hydrogen plant and the hydrocracker/fractionator complex are also expensive plants and along with the main hydrotreater account for almost half of the total capital investment. Total capital costs including off-sites and specified tankage, were \$878.6 million for maximum JP-4 production; \$862.5 million for maximum JP-8 production; and \$804.1 million for the JP-4 and other fuels case.

Using a proprietary in-house investment guidelines evaluation computer program, we calculated the total product costs to include both manufacturing and adjusted crude costs. (Note: Adjusted crude cost is defined as the cost of a barrel of crude multiplied by the ratio of total raw shale oil in to total liquid products out.) The results, which are presented in Table 51, were \$1.22/gal for the maximum JP-4 production; \$1.24/gal for the maximum JP-8 production; and \$1.19/gal for the JP-4 and other fuels case. Manufacturing costs were highest for the maximum JP-8 case, at \$10.38/Bbl of product; intermediate for the maximum JP-4 case, at \$10.30/Bbl of product; and lowest for the JP-4 and other fuels case at \$9.94/Bbl of product, as would be expected.

Yields, costs, and thermal efficiencies are summarized for the three optimized processing routes in Table 52. Based on total energy input to the refinery (crude, fuel, and utilities converted to FOE), 94.4 volume % jet fuel is produced when maximizing JP-4; 54.3 volume % jet fuel when maximizing JP-8; and 79.1 volume % jet fuel in the JP-4 plus other fuels case. Overall thermal efficiencies range from 80.7 to 86.7%. Plant

investments for the three shale oil refineries are between 8041 and 8786 \$/SDB of capacity. The plant investment for a conventional petroleum fuels refinery of similar capacity is less than half of the above figures.

Schematic flow diagrams for the three optimized processing schemes are presented in Figures 17 through 19.

## SECTION VI

### ANALYSIS OF DATA

The effect of changes in some of the economic variables given by the USAF guidelines would have on product fuel cost were examined. In particular we examined the effect of interest rate of return (IRR), raw shale oil prices, changes in required capital investment, annual interest rates, and percent of finance equity on product prices.

The sensitivity of product fuel cost to changes in the investment rate of return (IRR) is shown in Figure 20 and tabulated in Table 53. Due to the uncertainties associated with investing in a new technology, we feel that an IRR of at least 20% would be needed to attract capital. This change in IRR from 15 to 20% would increase the product fuel cost by an additional 9¢/gal.

The sensitivity of fuel cost to changes in the price of raw shale oil is presented in Figure 21 and Table 54. Changes in the price of raw shale oil has a significant impact on the product fuel cost. An increase in the cost of raw shale oil from \$40 to \$45/bbl would cause the product fuel cost to rise an additional 12¢/gal.

The effect of changes in capital investment on product fuel cost is summarized in Figure 22 and Table 55. A contingency was not included in the Phase IV economic evaluation. However, we recommend a contingency of at least 25% for new technology energy process plants such as a raw shale oil upgrading facility. An additional cost of 4.5¢/gal of product

results from the inclusion of a 25% contingency factor for a possible increase in capital investment.

The effect of changes in the annual interest rate of working capital on product fuel cost is shown in Figure 23 and Table 56. Since this interest charge is only associated with working capital and not plant investment, the product fuel cost is not overly sensitive to changes in the annual interest rate. Increasing the annual interest rate from 15% to 20% adds 0.4¢/gal to the product cost.

An increase in utilities cost by 25%, results in an additional increase of the product cost by 1.5¢/gal for the maximum JP-4 and JP-8 cases, and by 1.4¢/gal for the JP-4 and other fuels case.

Examining another scenario, where 100% of the capital investment would be borrowed at a 15% annual interest rate, we found that the product prices increase by 9.5¢/gal of product fuel. This might be the case where a federal loan guarantee could be obtained. Borrowing the investment capital at 20% interest rate results in an increase of fuel prices by 13¢/gal. A summary of these results is presented in Table 57.

## SECTION VII

### CONCLUSIONS

1. LP results showed that the optimal processing scheme for maximizing JP-4 and JP-8 was with the raw shale oil hydrotreater operating at 2200 ppm total nitrogen in the effluent, and for JP-4 and other fuels was with the raw shale oil hydrotreater operating at 6400 ppm total nitrogen in the effluent.
2. Total product yields as the volume percent of total refinery input (crude, fuel, and utilities converted to FOE BPSD) for maximum JP-4, maximum JP-8, and JP-4 and other fuels were 97.4, 95.5 and 100.7 respectively.
3. Total refinery hydrogen consumption was 2584, 2363, and 1960 SCF/Bbl of raw shale oil charged to the process units for the three cases stated above.
4. Overall refinery thermal efficiencies were 81.5, 80.7 and 86.7 respectively for the three cases stated above.
5. Economics were developed for a 100,000 BPSD refinery using a first quarter 1981 cost base and \$40 per Bbl for raw shale oil. Total product cost varied from \$1.19 to \$1.24 per gallon, depending on the refinery product slate.

6. Sensitivity analysis showed that product price was sensitive to the following in the order presented:

- ° Raw shale oil prices
- ° Discounted cash flow (IRR)
- ° Variations in capital investment
- ° Annual interest rate on working capital

7. Financing 100% of capital investment at 15% annual interest rate increased product prices by 9.5¢/gallon.

SECTION VIII  
RECOMMENDATIONS

It is recommend that:

1. Additional effort be expended to develop a safe method for arsenic disposal. The spent catalyst from the guard reactor will have a high arsenic content at the end of its useful life. Disposal or regeneration of the catalyst may present problems.
2. A determination be made whether or not raw shale oil is suitable for use as a fuel to furnaces for manufacturing hydrogen by steam reforming.
3. Continuous HCl extraction should be carried out in order to confirm or modify batch data. Continuous HCl extraction and recovery processes have not been demonstrated. Large scale runs were inhibited due to the lack of suitable continuous equipment.
4. Additional HCl extraction work should be carried out to determine how to control and minimize residual chloride concentrations in the raffinate and extract phases. The chloride content in the HCl raffinate varied randomly over a wide range from run to run, from 100 to 800 ppm Cl.



5. Pilot plant hydrocracking with extinction recycle be demonstrated to firm up data generated by Sun Tech's Hydrocracking Math Model. Extinction recycle of hydrocracked bottoms was not demonstrated in the pilot plant. It probably can be done to maximize JP-4, but not for JP-8 due to build up of aromatics and wax in the recycle stream. Yields and operating conditions were obtained from Sun Tech's Hydrocracking Math Model.
6. Market values for products be used in the LP model to give a more realistic optimized processing scheme. The LP optimization program is price driven and the optimized results are only as realistic as the economics used.
7. A minimum contingency of 25% be used in the economic evaluations of new technology energy process plants such as a shale oil upgrading facility. A contingency was not included in the Phase IV economics.

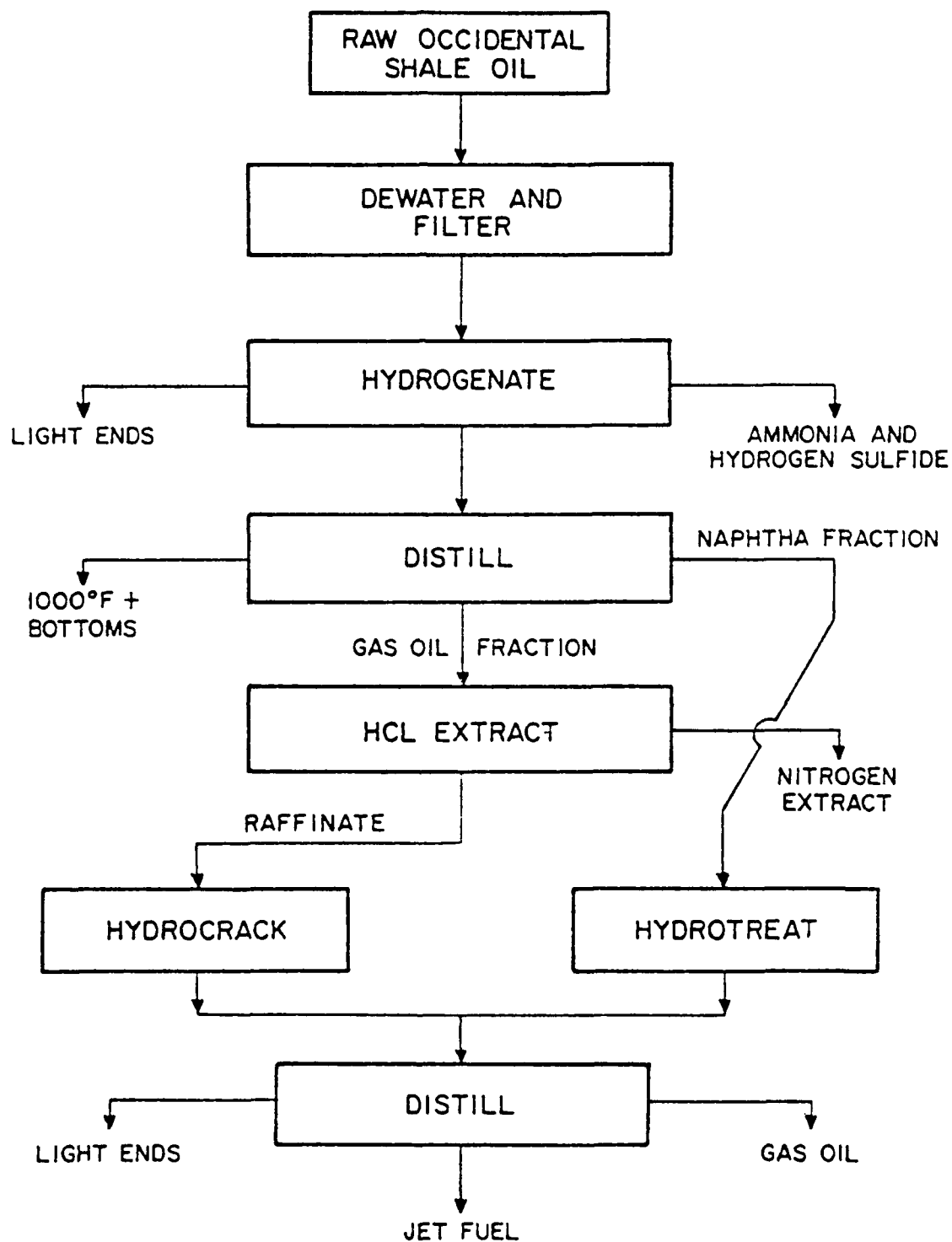


FIGURE 1  
BLOCK FLOW DIAGRAM FOR PREPARING  
JET FUELS FROM OCCIDENTAL SHALE OIL

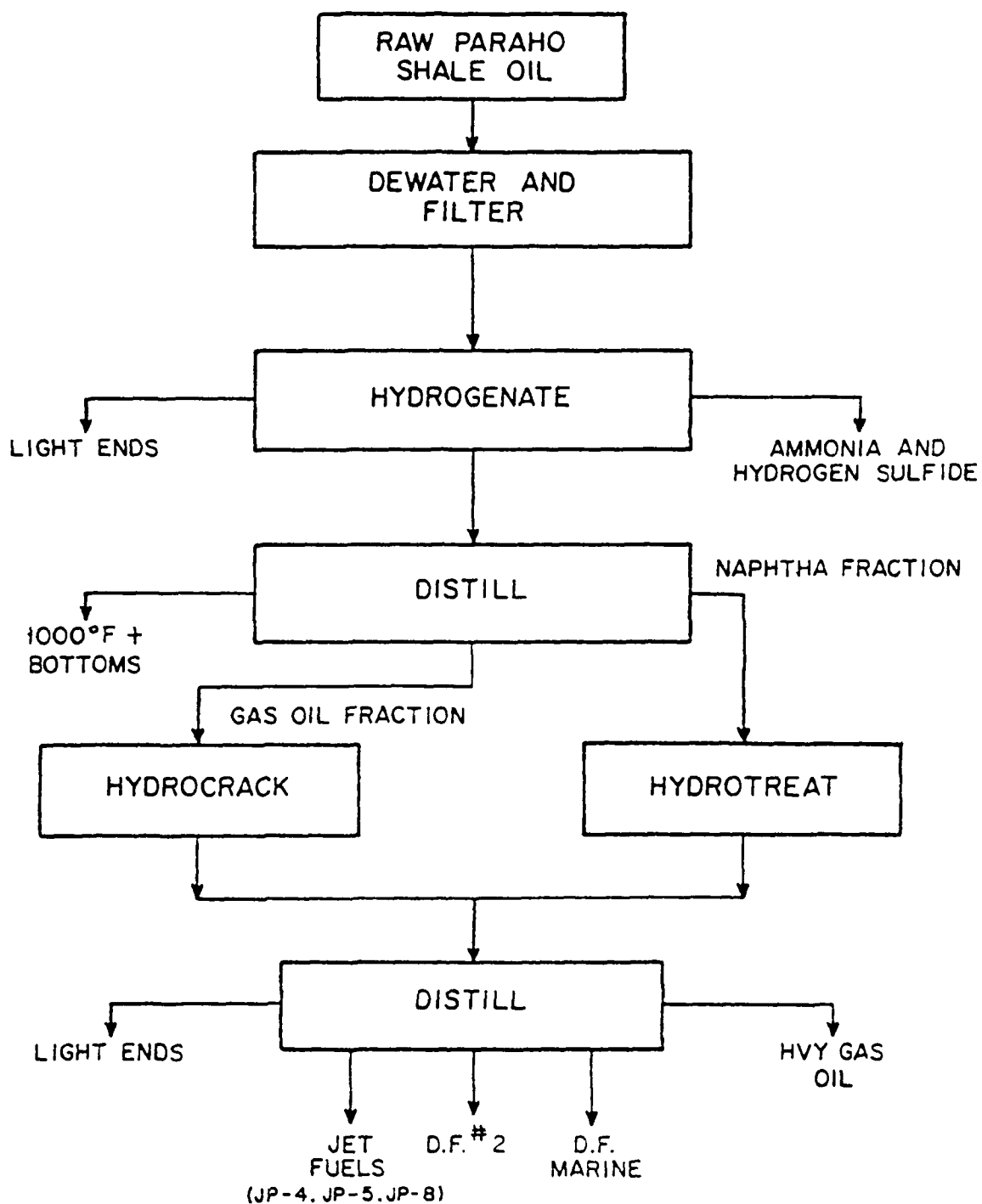


FIGURE 2  
BLOCK FLOW DIAGRAM FOR PREPARING  
MILITARY FUELS FROM PARAHO SHALE OIL

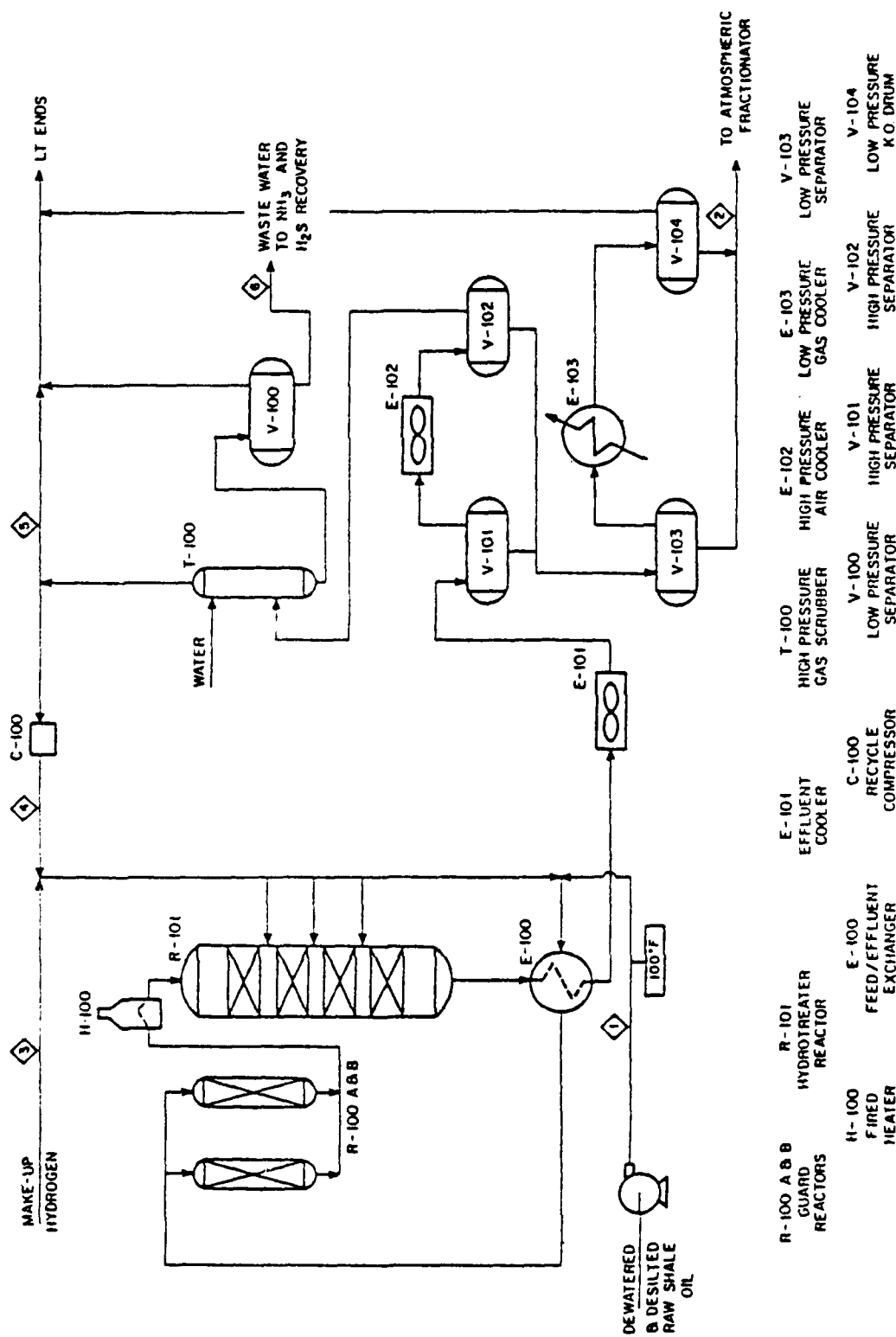


FIGURE 3

# RAW SHALE OIL HYDROTREATER SIMPLIFIED FLOW DIAGRAM

R-100 A & B	R-101	E-101	C-100	E-102	E-103	V-103	V-104
GUARD REACTORS	HYDROTREATER REACTOR	EFFLUENT COOLER	RECYCLE COMPRESSOR	HIGH PRESSURE AIR COOLER	LOW PRESSURE GAS COOLER	LOW PRESSURE SEPARATOR	LOW PRESSURE K.O. DRUM
H-100	E-100	T-100	V-101	V-102	V-103	V-104	
FIRIED HEATER	FEED/EFFLUENT EXCHANGER	HIGH PRESSURE GAS SCRUBBER	HIGH PRESSURE SEPARATOR	HIGH PRESSURE SEPARATOR	LOW PRESSURE SEPARATOR	LOW PRESSURE SEPARATOR	



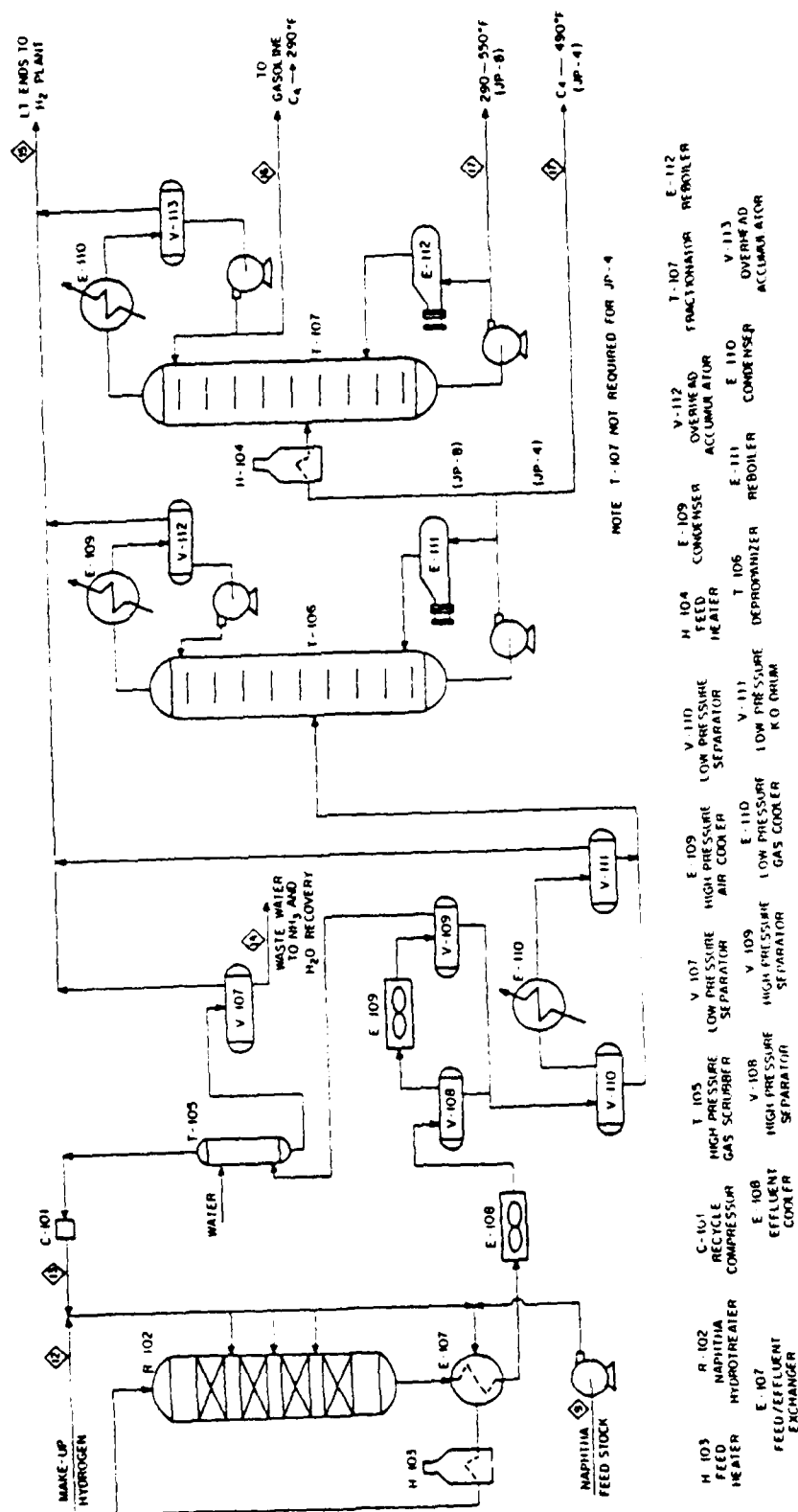


FIGURE 5  
NAPHTHA HYDROTREATER AND FRACTIONATOR  
SIMPLIFIED FLOW DIAGRAM

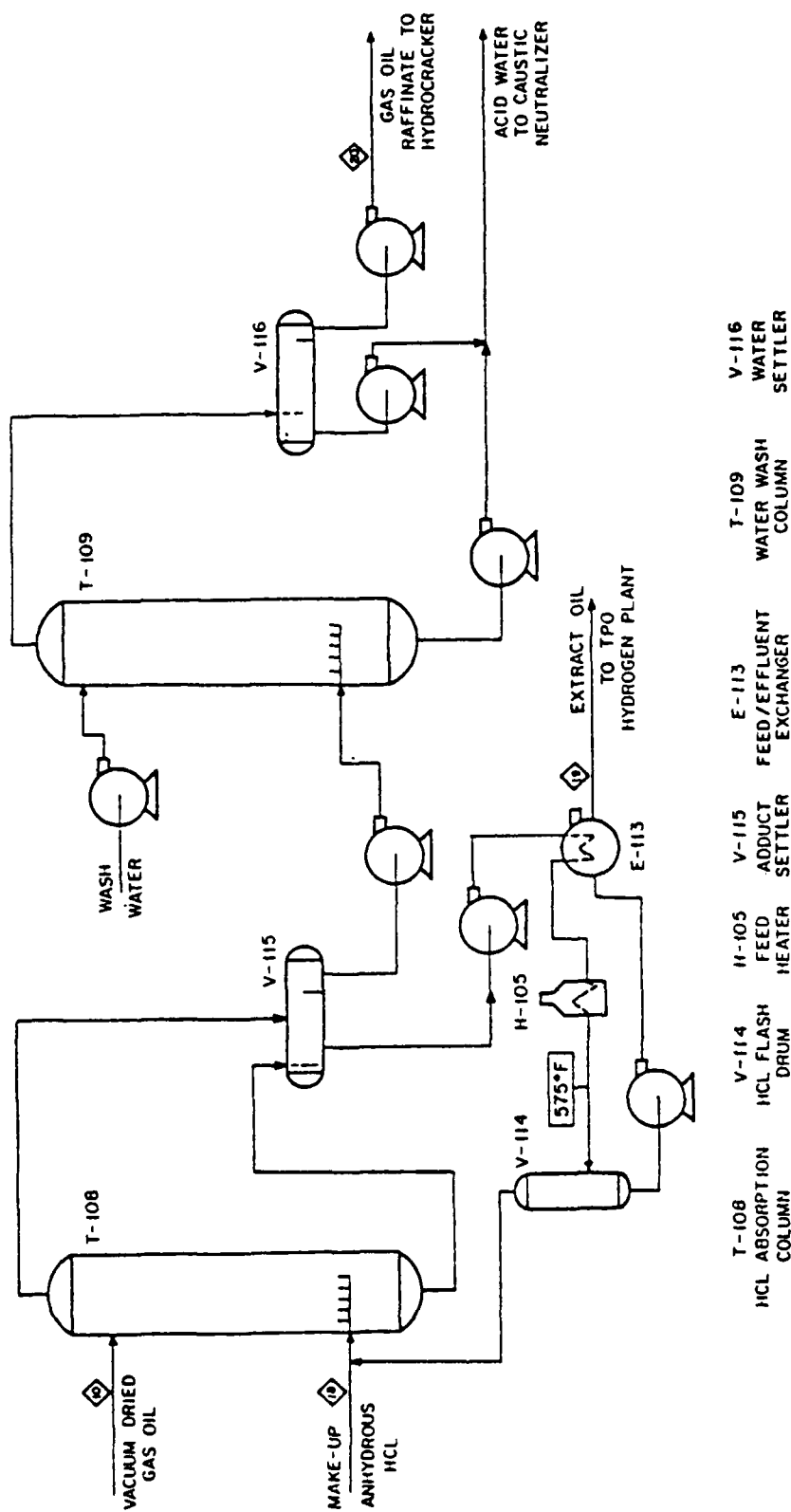


FIGURE 6  
ANHYDROUS HCL EXTRACTION PLANT  
SIMPLIFIED FLOW DIAGRAM





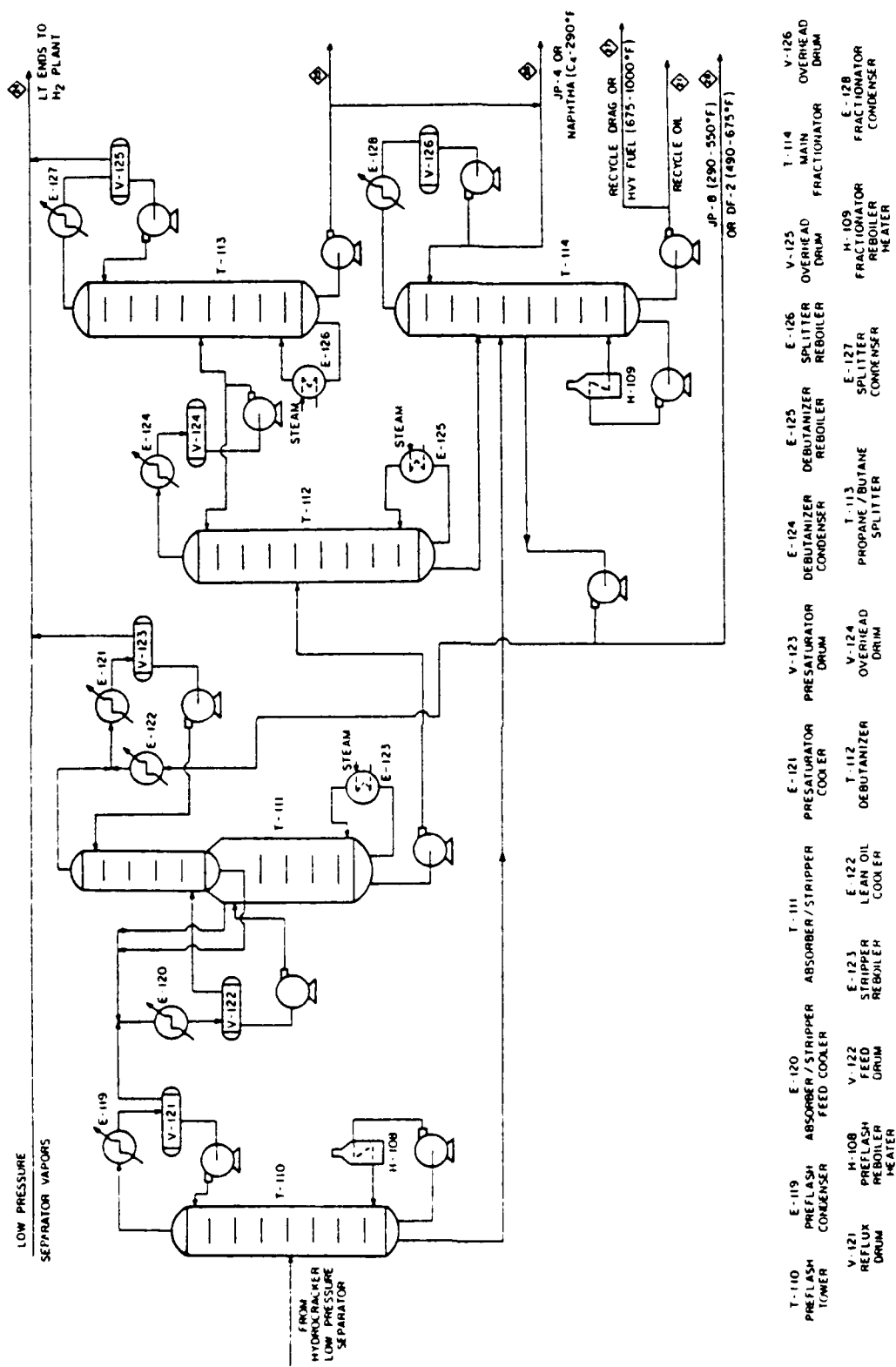


FIGURE 8  
HYDROCRACKER FRACTIONATOR  
SIMPLIFIED FLOW DIAGRAM

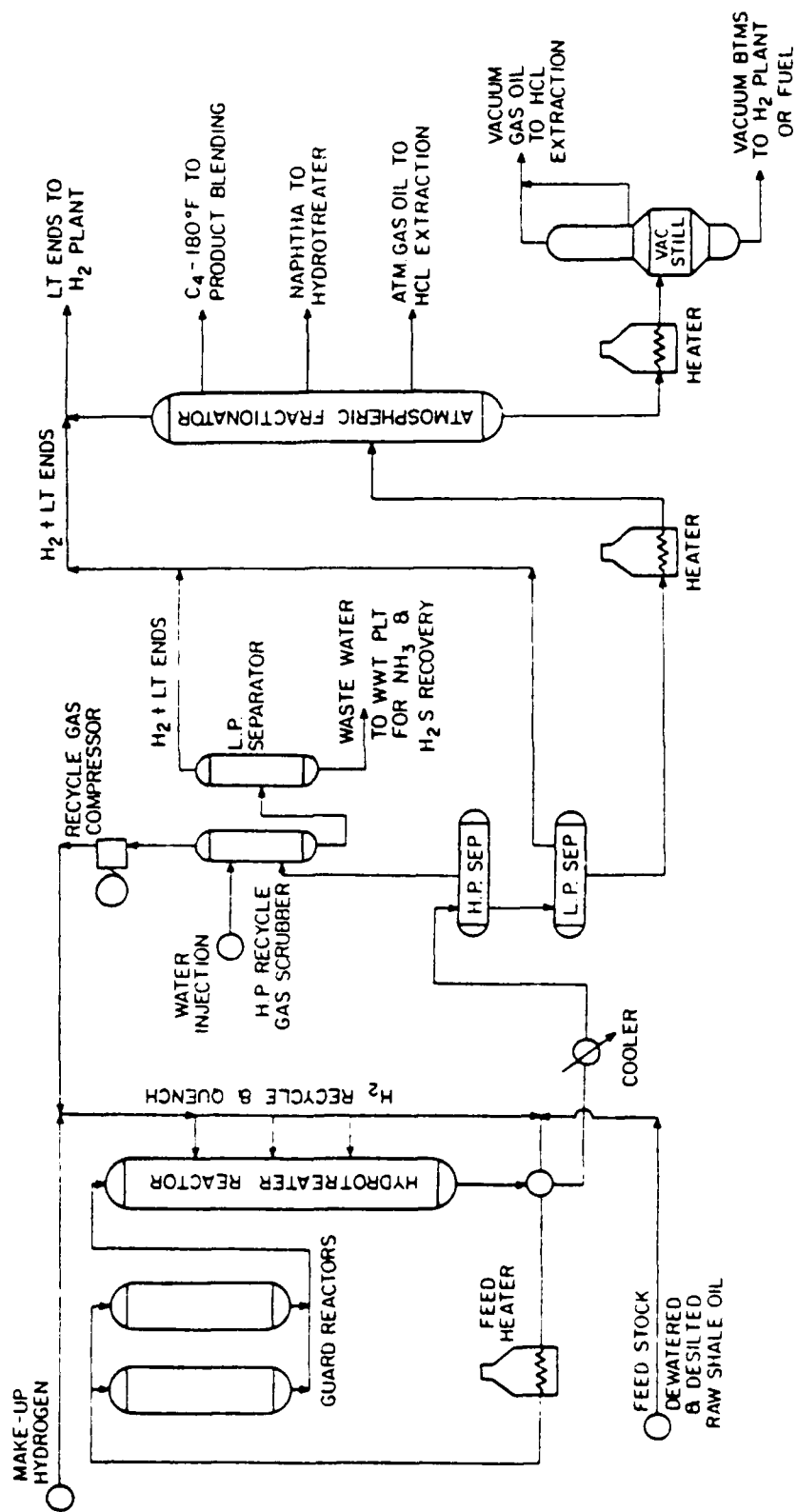


FIGURE 9  
SIMPLIFIED FLOW DIAGRAM OF RAW SHALE OIL HYDROTREATER  
AND DISTILLATION PLANTS

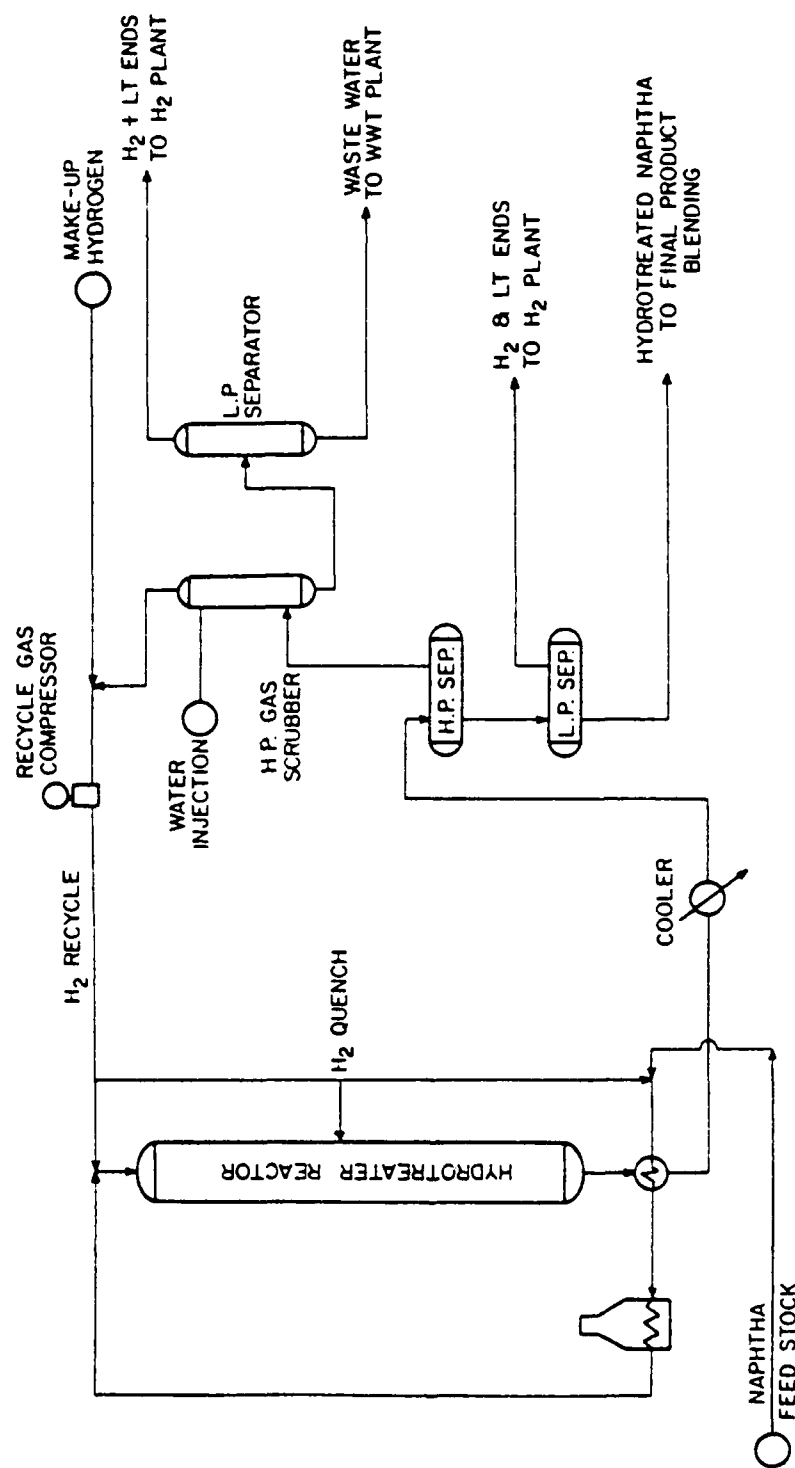


FIGURE 10  
SCHEMATIC FLOW DIAGRAM OF NAPHTHA HYDROTREATER

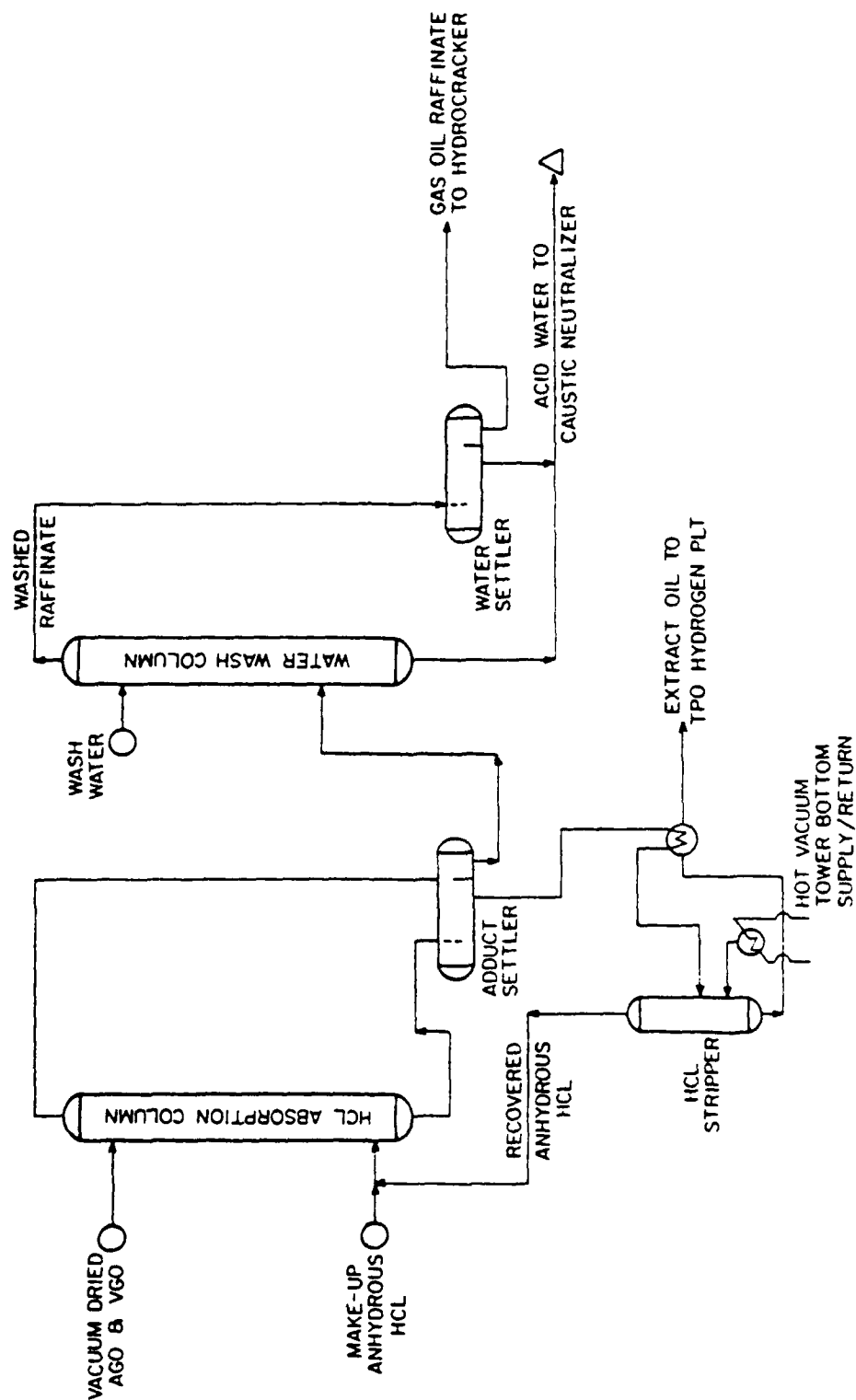


FIGURE 11  
SCHEMATIC FLOW DIAGRAM OF ANHYDROUS  
HCL EXTRACTION PLANT



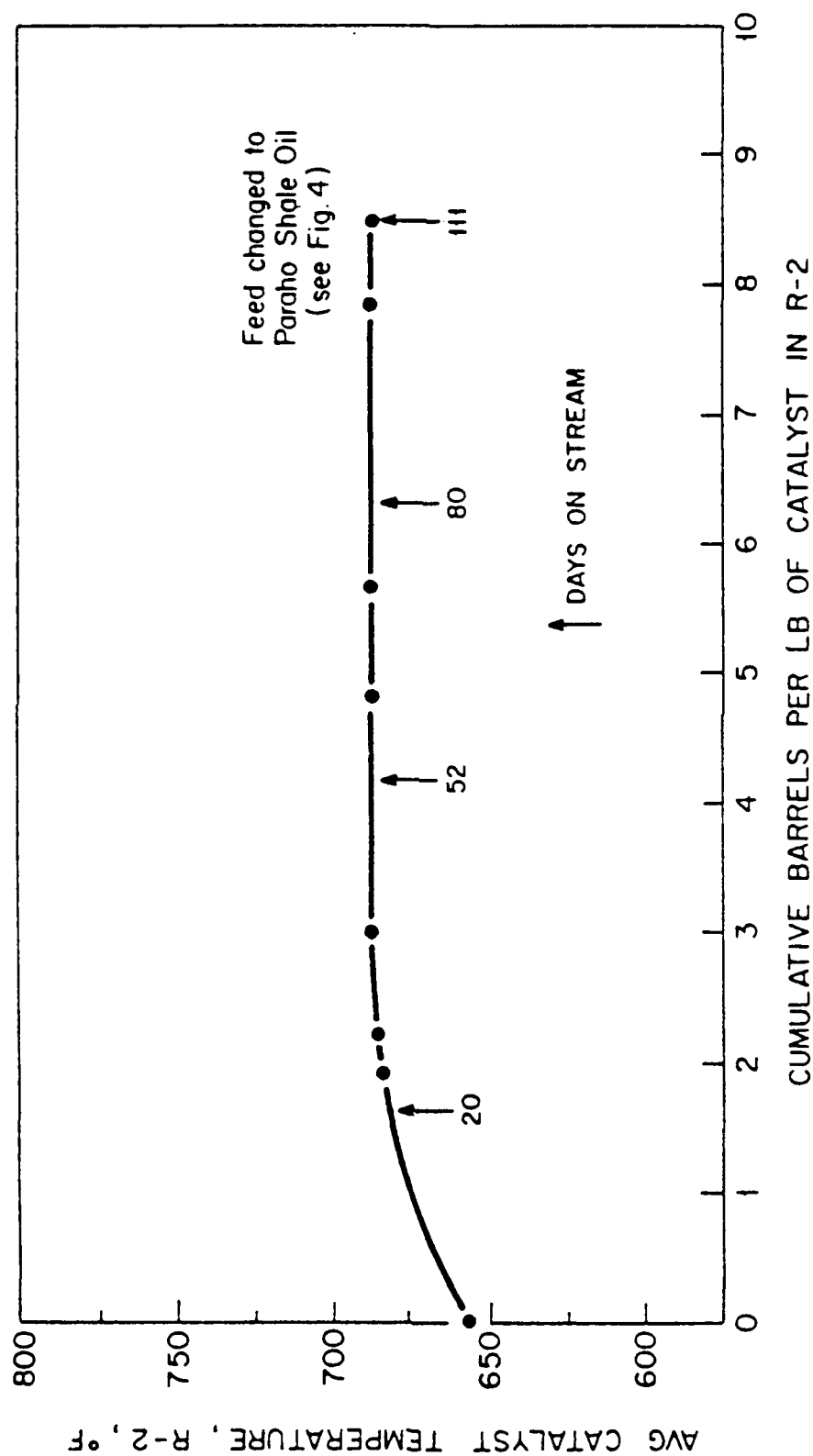


FIGURE 13  
 CATALYST LIFE TEST FOR HYDROTREATING  
 WHOLE OCCIDENTAL SHALE OIL  
 (TO 5,000 ppm  $N_T$  PRODUCT)

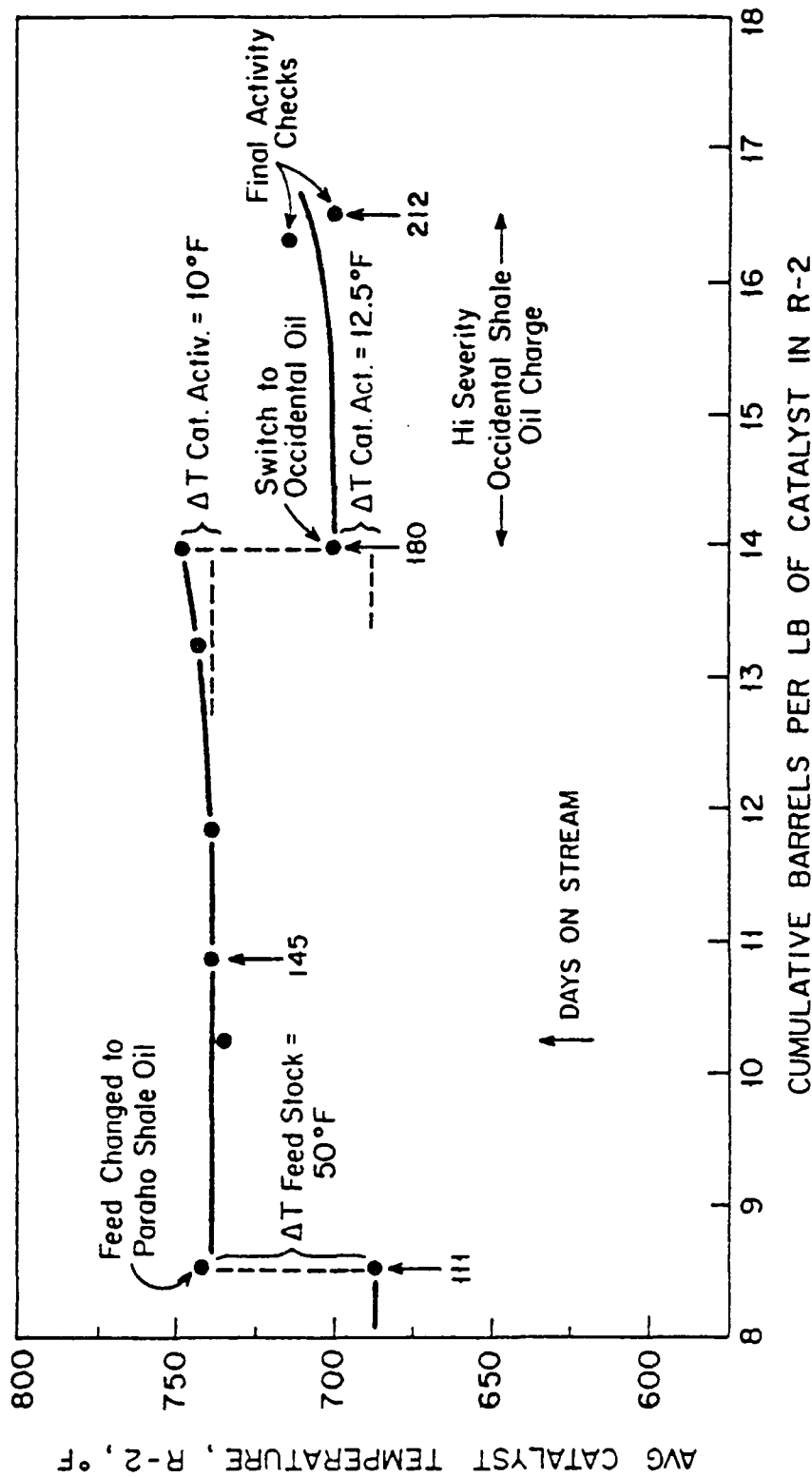
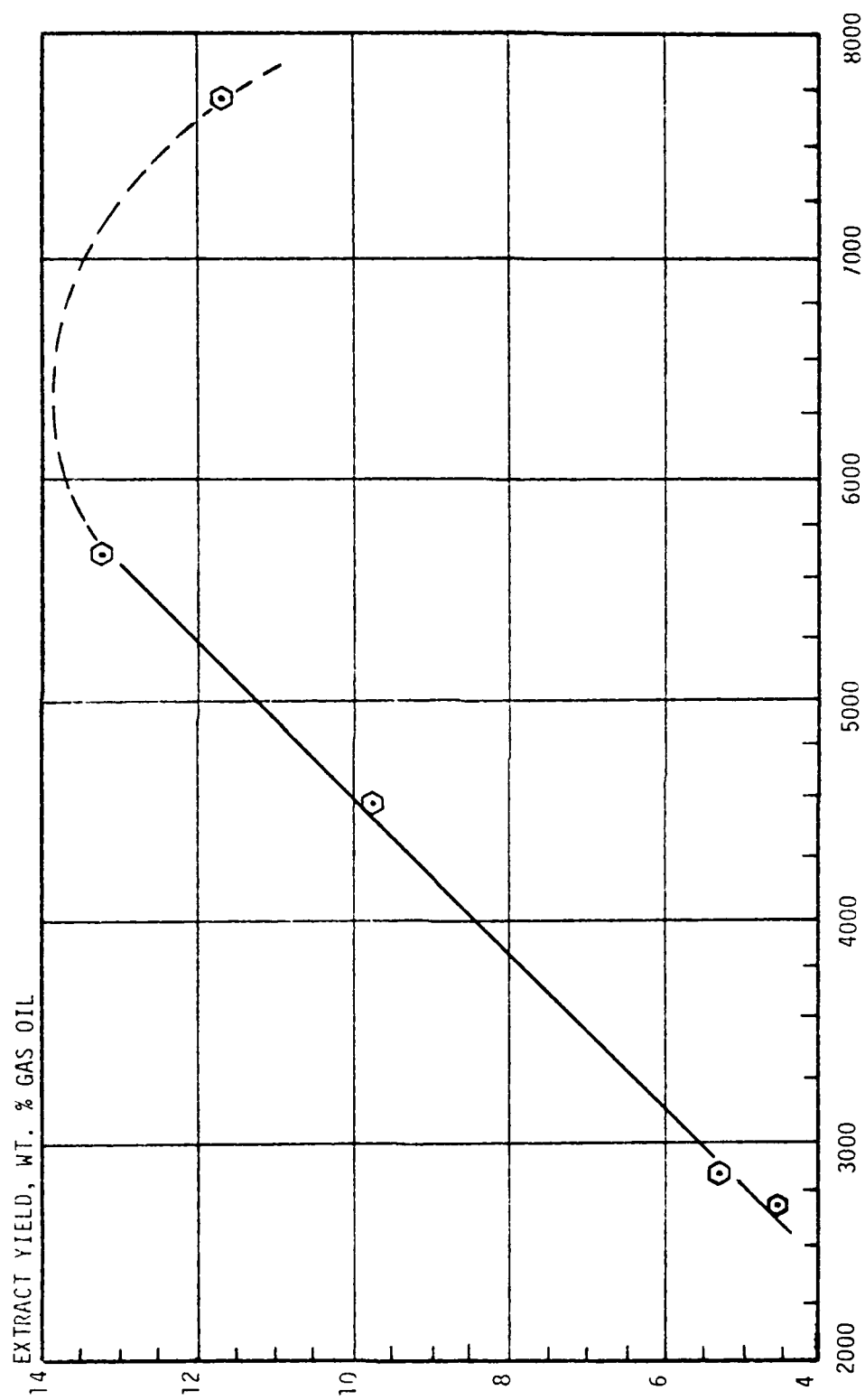


FIGURE 14

CATALYST LIFE TEST FOR HYDROTREATING  
WHOLE PARAHO SHALE OIL  
(TO 5,000 ppm  $N_T$  PRODUCT)







N<sub>T</sub> CONTENT IN 535 - 1,000°F FRACTION, PPM

FIGURE 16  
EFFECT OF NITROGEN CONTENT IN GAS OIL  
FRACTION ON HCL EXTRACT YIELD

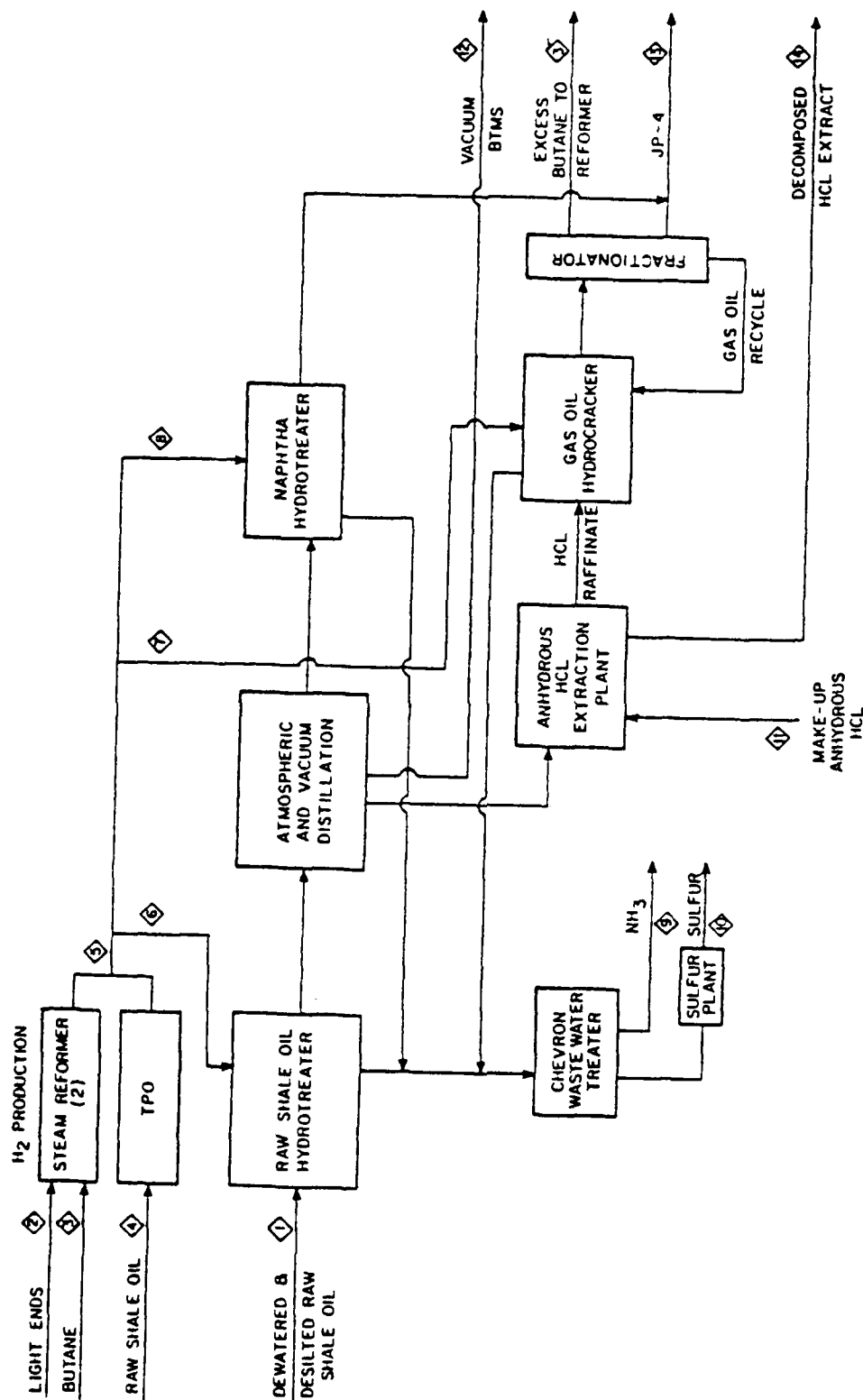


FIGURE 17  
SCHEMATIC FLOW DIAGRAM FOR REFINING RAW SHALE OIL  
USING ANHYDROUS HCL EXTRACTION  
JP-4 OPERATION



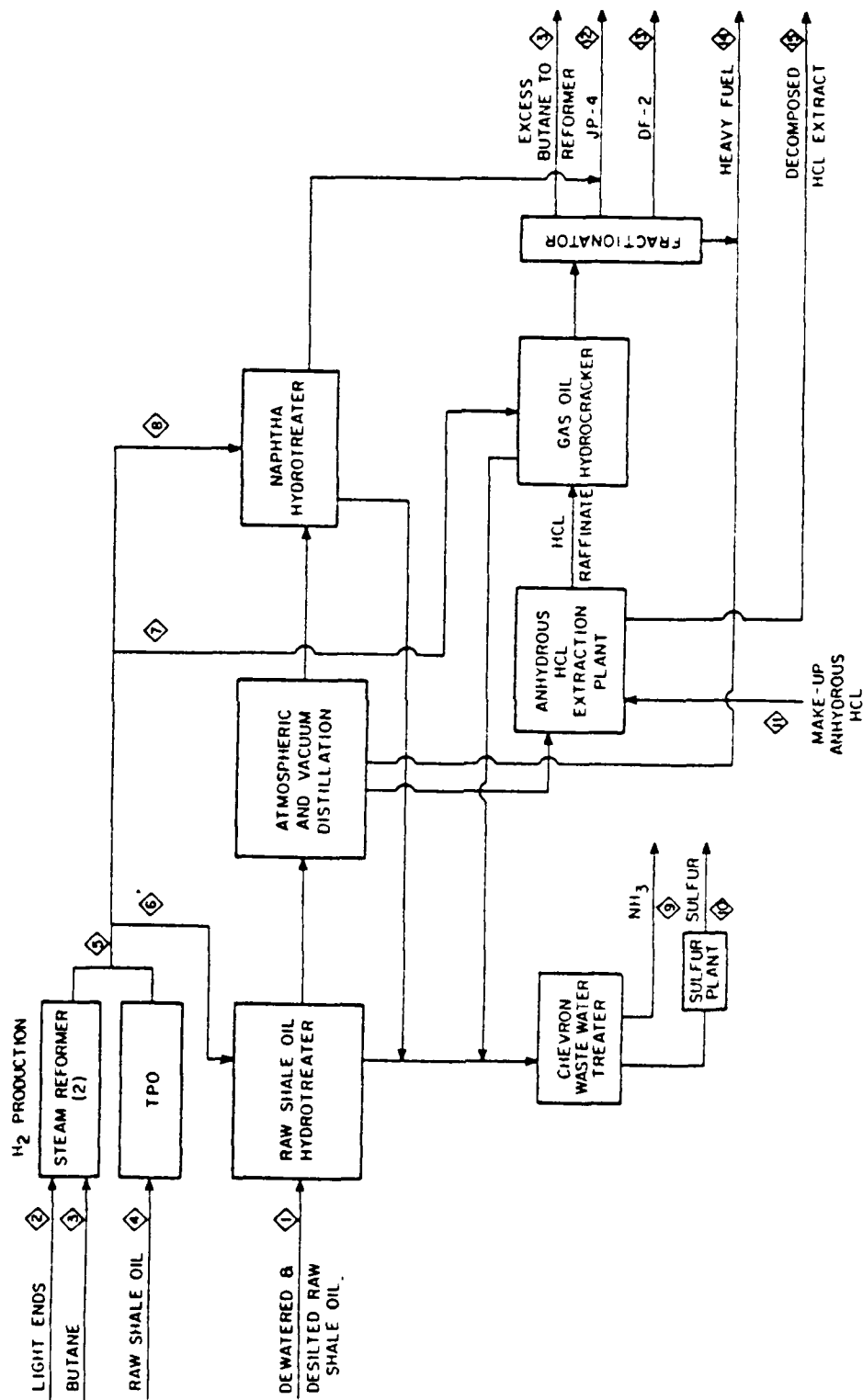


FIGURE 19  
SCHEMATIC FLOW DIAGRAM FOR REFINING RAW SHALE OIL  
USING ANHYDROUS HCL EXTRACTION  
JP-4 AND OTHER FUELS OPERATION

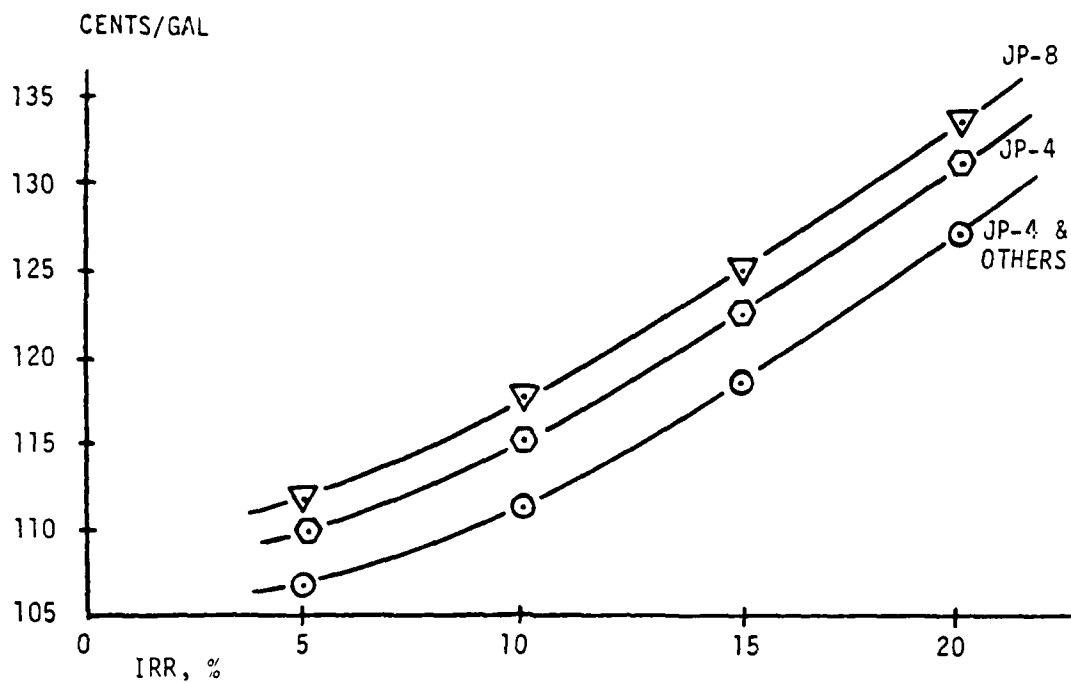


FIGURE 20  
SENSITIVITY OF FUEL COST TO CHANGES  
IN INTEREST RATE OF RETURN (IRR)

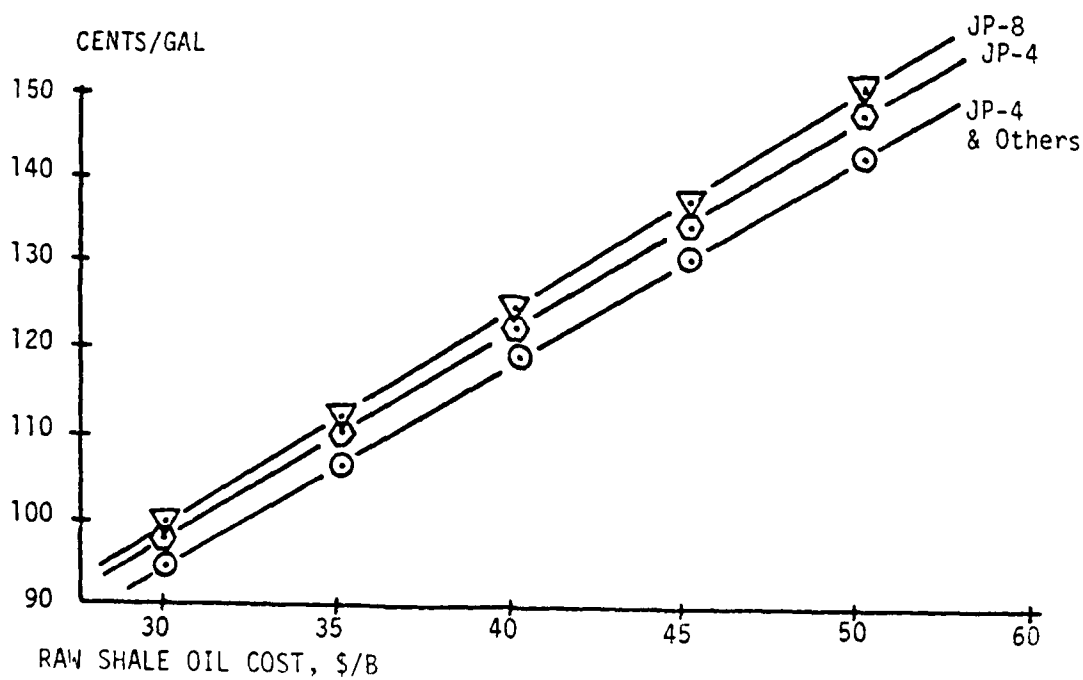


FIGURE 21  
SENSITIVITY OF FUEL COST TO  
CHANGES IN PRICE OF RAW SHALE OIL

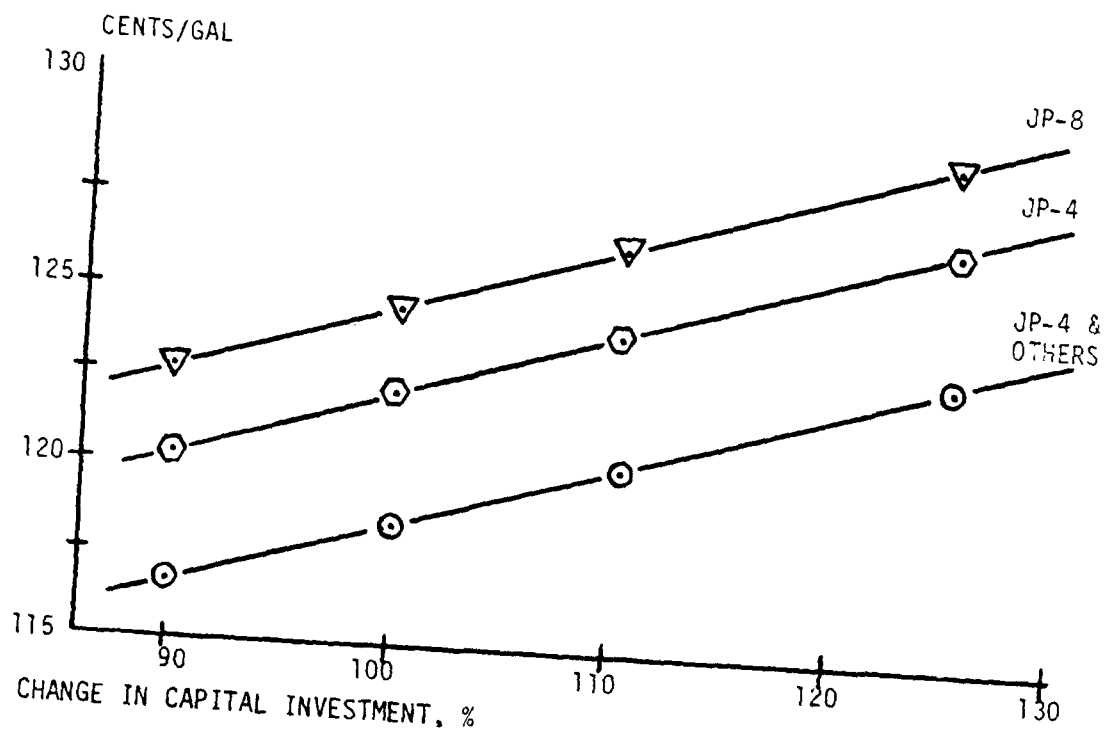


FIGURE 22  
SENSITIVITY OF FUEL COST TO  
CHANGES IN CAPITAL INVESTMENT

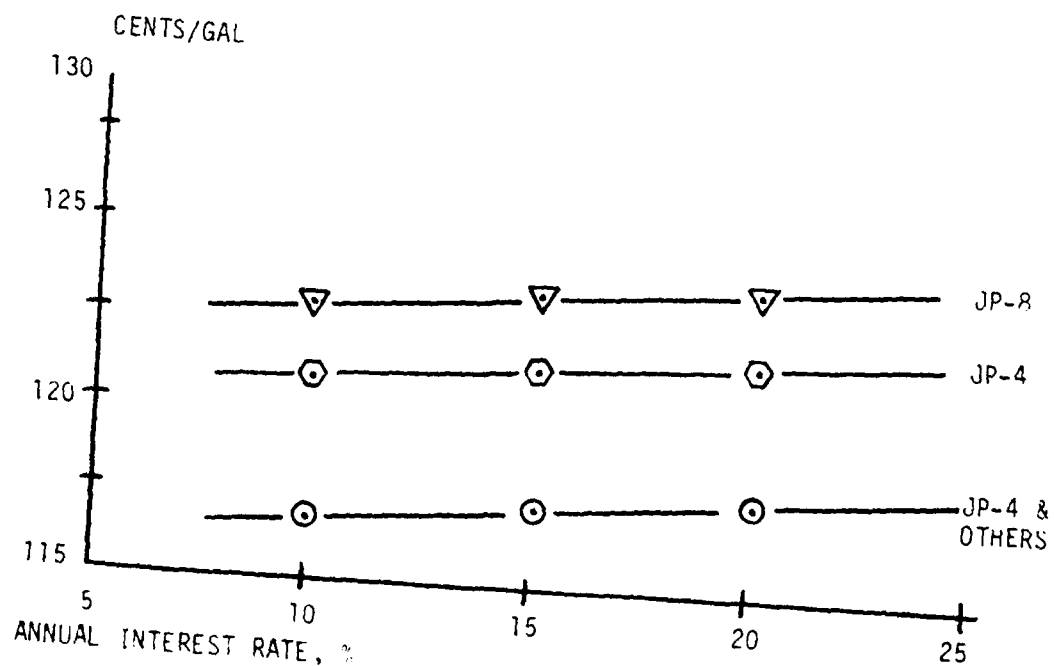


FIGURE 23  
SENSITIVITY OF FUEL COST TO CHANGES IN  
ANNUAL INTEREST RATE OF WORKING CAPITAL

TABLE 1

INSPECTIONS AND ANALYSES OF RAW SHALE OIL

Raw Shale Oil	<u>Paraho</u>	<u>Occidental</u>
Inspection Data		
API @ 60°F	20.6	23.0
Specific Gravity 60/60	0.9303	0.9160
Viscosities, KV		
@ 100°F, cs	60	32.3
@ 210°F, cs	5.38	4.82
Distillation, °F		
IBP	D1160	D2887
10 Vol. %	133	296
30   "	508	459
50   "	687	558
70   "	798	649
90   "	918	768
FBP	1057	876
FBP	1065/95%	1071
Ramsbottom Carbon Res., Wt.%	1.4	-
Asphaltenes, Wt.%	-	2.4
Chemical Composition Data, Wt.%		
Carbon	83.83	84.82
Hydrogen	11.72	12.04
Oxygen	1.31	1.18
Nitrogen (Total)	2.13	1.46
(Basic)	1.31	0.81
Sulfur	0.75	0.62
Iron, ppm	90	NA
Arsenic, ppm	34	33

TABLE 2  
ESTIMATED HYDROGEN YIELDS FROM VARIOUS FEEDSTOCKS  
BASED ON TEXACO PARTIAL OXIDATION PROCESS

Inspections	Decomposed HCl Extract		1000°F+ Bottoms From		Raw Shale Oil			
	From Paraho Shale Oil		From Occid. Shale Oil		Hydrotreated Occidental Shale Oil		Paraho	
	Texaco	Sun Tech	Sun Tech	Sun Tech	Sun Tech	Sun Tech	Sun Tech	Sun Tech
Sp. Gr., 60/60	0.9459	0.9458	0.9626	0.9593	0.9159	0.9303	0.9159	0.8939
API Gravity	18.1	18.1	15.5	16.0	23.0	20.6	23.0	26.8
Lbs/Bbl	330.834	336.672	335.538	335.538	320.334	325.374	320.334	312.606
Composition, wt. %								
Carbon	85.86	85.86	83.88	86.39	84.82	83.83	84.82	84.48
Hydrogen	11.14	11.14	10.80	12.33	12.04	11.72	12.04	11.69
Nitrogen	2.57	2.57	3.45	0.79	1.46	2.13	1.46	1.66
Req'd Oxygen, Lbs/Bbl	368.13	369.40	365.99	375.67	352.13	353.50	352.13	342.26
Req'd Steam, Lbs/Bbl	135.6	135.6	138.04	137.57	131.34	133.4	131.34	128.17
Hydrogen, 100%								
SCF/Bbl	15,508	15,386	15,406	15,814	14,823	14,880	14,823	14,407
Hydrogen Purity,								
Mol %	97.89	97.79	97.54	98.18	98.02	97.85	98.02	97.99



TABLE 3

FEED AND UTILITY REQUIREMENTS FOR TEXACO PARTIAL OXIDATION PROCESSBasis: 1000 SCF H<sub>2</sub> + COFEED AND UTILITY REQUIREMENTS

Feedstock (Nitrogen Extract)	21.3 lbs.
Oxygen (100% Basis)	23.7 lbs.
Steam (Superheated to 800°F)	8.74 lbs.
Ratio O <sub>2</sub> /(CO + H <sub>2</sub> )	0.27
Electric Power	1.7 Kwh
Cooling Water (35° ΔT)	95 gal.
Boiler Feed Water	91.4 lbs.

PRODUCT GASESMol % Dry Basis

Carbon Monoxide	48.40
Hydrogen	46.31
Carbon Dioxide	4.30
Methane	0.33
Argon	0.11
Nitrogen	0.55
Hydrogen Sulfide	22 ppm
Carbonyl Sulfide	<u>1 ppm</u>

TOTAL 100.00

Unreacted Carbon, lbs./hr. 0.36

TABLE 4

OPERATING CONDITIONS FOR PROCESSING WHOLE OCCIDENTAL SHALE OIL

## BASIS:

CHARGE RATE: 100,000 BPSD (90,000 BPCD) Raw Occidental Shale Oil

OPERATING FACTOR: 0.90

CATALYSTS: NiMo on Spherical Alumina (R-1)  
NiMo on Alumina (R-2)REACTOR OPERATING CONDITIONSCASE

Total Nitrogen in liquid effluent, ppm	<u>2200</u>	<u>5000</u>	<u>6400</u>
LHSV, V/hr/V, R-1	1.0	1.0	1.0
R-2	1.0	1.0	1.0
Catalyst Life, months			
R-1	6	6	6
R-2	18	24	30
Avg. Catalyst Temp, °F			
R-1	625	625	625
R-2	730	690	665
Pressure, Total PSIA	1615	1615	1615
H <sub>2</sub> PP	1520	1520	1520
Recycle Gas Rate, SCF/B	4000	4000	4000
Hydrogen Consumption, SCF/B			
Chemical	1320	1100	900
Dissolved	150	150	150
Bleed	100	100	100
Total to Hydrotreater	1570	1350	1150
Product Data			
Total Nitrogen, ppm	2200	5000	6400
Sulfur, ppm	170	140	306
C <sub>4</sub> + Yield, Vol.% Feed	103.90	103.55	102.97

TABLE 5

MATERIAL BALANCE SUMMARY FOR MAIN HYDROTREATER  
AND DISTILLATION UNITS (2200 ppm N<sub>T</sub>)

---

## BASIS:

100,000 BPSD Raw Occidental Shale Oil

157 x 10<sup>6</sup> SCF Hydrogen PSD (132 x 10<sup>6</sup> SCF H<sub>2</sub> Chemically Consumed  
PSD)

Liquid Effluent Treated to 2200 ppm Total Nitrogen

<u>PRODUCTS</u>	<u>JP-4</u>	<u>JP-8</u>
Ammonia, STSD	242	242
Hydrogen Sulfide, Sulfur Eq. STSD	112	112
Unreacted H <sub>2</sub> , SCF x 10 <sup>6</sup> SCF PSD	25.0	25.0
C <sub>1</sub> -C <sub>3</sub> Gases, Lbs. PSD	376,750	376,750
<u>FRACTION TBP CUT POINTS</u>		
C <sub>4</sub> -180°F, BPSD	738	
180-490°F, BPSD	27,132	
490-1000°F, BPSD	71,904	
C <sub>4</sub> -290°F, BPSD		5,380
290-550°F, BPSD		24,450
550-1000°F, BPSD		69,944
1000°F+ Bottoms, BPSD	4,126	4,126
TOTAL LIQUIDS, BPSD	103,900	103,900

TABLE 6

PRODUCT INSPECTIONS ON STREAMS FROM MAIN  
HYDROTREATER DISTILLATION UNIT (2200 ppm NT)

<u>INSPECTIONS</u>	<u>180-490°F</u>	<u>490-1000°F</u>	<u>290-550°F</u>	<u>550-1000°F</u>	<u>1000°F+ BTMS.</u>
API Gravity @ 60°F	42.5	29.6	42.0	29.4	25.3
Total Nitrogen, ppm	810	2400	863	2422	7048
Sulfur, ppm	55	107	57	108	415

TABLE 7

MATERIAL BALANCE SUMMARY FOR MAIN HYDROTREATER  
AND DISTILLATION UNITS (5000 ppm N<sub>T</sub>)

---

## BASIS:

100,000 BPSD Raw Occidental Shale Oil

135 x 10<sup>6</sup> SCF Hydrogen PSD (110 x 10<sup>6</sup> SCF H<sub>2</sub> Chemically Consumed  
PSD)

Liquid Effluent Treated to 5000 ppm Total Nitrogen

<u>PRODUCTS</u>	<u>JP-4</u>	<u>JP-8</u>
Ammonia, STSD	187	187
Hydrogen Sulfide, Sulfur Eq. STSD	110	110
Unreacted H <sub>2</sub> , SCF x 10 <sup>6</sup> SCF PSD	25.0	25.0
C <sub>1</sub> -C <sub>3</sub> Gases, Lbs. PSD	385,294	385,294
<u>FRACTION TBP CUT POINTS</u>		
C <sub>4</sub> -180°F, BPSD	2,116	
180-490°F, BPSD	24,141	
490-1000°F, BPSD	73,133	
C <sub>4</sub> -290°F, BPSD		4,550
290-550°F, BPSD		25,561
550-1000°F, BPSD		69,279
1000°F+ Bottoms, BPSD	4,159	4,159
TOTAL LIQUIDS, BPSD	103,549	103,549

TABLE 8  
PRODUCT INSPECTIONS ON STREAMS FROM MAIN  
HYDROTREATER DISTILLATION UNIT (5000 ppm N<sub>T</sub>)

INSPECTIONS	180-490°F	490-1000°F	C <sub>4</sub> -290°F	290-550°F	550-1000°F	1000°F+ BTMS.
API Gravity @ 60°F	41.5	28.9	71.0	40.6	28.6	16.0
Distillation, °F						
I/10	180/290	490/605	50/145	290/360	550/606	NA
30/50	405/441	661/734	183/202	433/458	671/744	--
70/95	468/486	817/965	217/252	480/526	820/970	--
EP	490	1000	290	550	1000	--
Aromatics, Wt. %	24 Vol. %	42	3 Vol. %	25	45	50
Olefins, Vol. %	3	--	--	3	--	--
Total Nitrogen, ppm	3260	4800	--	3480	5600	7900
Sulfur, ppm	65	140	--	80	150	1220

TABLE 9

MATERIAL BALANCE SUMMARY FOR MAIN HYDROTREATER  
AND DISTILLATION UNITS (6400 ppm N<sub>T</sub>)

---

## BASIS:

100,000 BPSD Raw Occidental Shale Oil

120 x 10<sup>6</sup> SCF Hydrogen PSD ( 95 x 10<sup>6</sup> SCF H<sub>2</sub> Chemically Consumed  
PSD)

Liquid Effluent Treated to 6400 ppm Total Nitrogen

<u>PRODUCTS</u>	<u>JP-4</u>	<u>JP-8</u>
Ammonia, STSD	162	162
Hydrogen Sulfide, Sulfur Eq. STSD	109	109
Unreacted H <sub>2</sub> , SCF x 10 <sup>6</sup> SCF PSD	25.0	25.0
C <sub>1</sub> -C <sub>3</sub> Gases, Lbs. PSD	324,365	324,365
<u>FRACTION TBP CUT POINTS</u>		
C <sub>4</sub> -180°F, BPSD	2,932	
180-490°F, BPSD	19,808	
490-1000°F, BPSD	76,031	
C <sub>4</sub> -290°F, BPSD		4,332
290-550°F, BPSD		26,011
550-1000°F, BPSD		68,428
1000°F+ Bottoms, BPSD	4,202	4,202
TOTAL LIQUIDS, BPSD	102,973	102,973

TABLE 10

PRODUCT INSPECTIONS ON STREAMS FROM MAIN  
HYDROTREATER DISTILLATION UNIT (6400 ppm N<sub>T</sub>)

<u>INSPECTIONS</u>	<u>180-490°F</u>	<u>490-1000°F</u>	<u>290-550°F</u>	<u>550-1000°F</u>	<u>1000°F+ BTMS.</u>
API Gravity @ 60°F	40.3	28.1	38.6	27.4	23.4
Total Nitrogen, ppm	3940	6887	4200	7104	8300
Sulfur, ppm	85	306	95	326	1245



TABLE 11

OPERATING CONDITIONS FOR NAPHTHA  
HYDROTREATER (2200 ppm N<sub>T</sub>)

---

Operator Factor: 0.90

Catalyst: NiMo on Alumina  
Catalyst Life: 2.5 Years

REACTOR OPERATING CONDITIONS:

	CASE	
	MAX. JP-4	MAX. JP-8
Feedstock TBP Boiling		
Range, °F	180-490	180-550
Total Nitrogen, ppm	810	863
LHSV, V/Hr/V	2.0	2.0
Avg. Catalyst Temp., °F	725	725
Total Pressure, psia	1500	1500
H <sub>2</sub> PP	1400	1400
Recycle Gas Rate, SCF/B	4000	4000
Hydrogen Consumption, SCF/B		
Chemical	230	240
Dissolved	50	50
Total to Hydrotreater	280	290
Product		
Total Nitrogen, ppm	8	8
Sulfur, ppm	2	2
C <sub>4</sub> + Yield, Vol.% Feed	101.50	100.60

TABLE 12

MATERIAL BALANCE SUMMARY FOR  
NAPHTHA HYDROTREATER (2200 ppm N<sub>T</sub>)

---

JP-4 OPERATION

Basis: 27,132 BPSD of 180-490°F Naphtha Fraction

7.60 x 10<sup>6</sup> SCF Hydrogen PSD (6.24 x 10<sup>6</sup> SCH H<sub>2</sub> Chemically Consumed)

PRODUCTS

Ammonia, STSD	3.8	<u>INSPECTIONS ON 180-490°F. CUT</u>		
Hydrogen Sulfide, Sulfur, Eq, STSD	0.2		<u>FEED</u>	<u>PRODUCT</u>
Unreacted H <sub>2</sub> x 10 <sup>6</sup> SCF PSD	1.36	API Gravity @ 60°F	42.5	42.9
C <sub>1</sub> -C <sub>3</sub> Gases, Lbs PSD	15,686	Total Nitrogen, ppm	810	8.0
C <sub>4</sub> -180°F, BPSD	1,357	Sulfur, ppm	55	2.0
180-490°F, BPSD	<u>26,182</u>			
TOTAL LIQUIDS, BPSD	27,539			

TABLE 13

MATERIAL BALANCE SUMMARY FOR  
NAPHTHA HYDROTREATER (2200 ppm N<sub>T</sub>)

---

JP-8 OPERATION

Basis: 29,092 BPSD of 180-550°F Naphtha Fraction

8.44 x 10<sup>6</sup> SCF Hydrogen PSD (6.98 x 10<sup>6</sup> SCH H<sub>2</sub> Chemically Consumed)

PRODUCTS

Ammonia, STSD	4.4	<u>INSPECTIONS ON 180-490°F. CUT</u>		
Hydrogen Sulfide, Sulfur, Eq, STSD	0.2		<u>FEED</u>	<u>PRODUCT</u>
Unreacted H <sub>2</sub> x 10 <sup>6</sup> SCF PSD	1.46	API Gravity @ 60°F	42.0	42.4
C <sub>1</sub> -C <sub>3</sub> Gases, Lbs PSD	16,819			
		Total Nitrogen, ppm	863	8.0
C <sub>4</sub> -180°F, BPSD	5,130			
290-550°F, BPSD	<u>24,427</u>	Sulfur, ppm	57	2.0
TOTAL LIQUIDS, BPSD	29,557			

TABLE 14

OPERATING CONDITIONS FOR  
NAPHTHA HYDROTREATER (5000 ppm N<sub>T</sub>)

---

Operator Factor: 0.90

Catalyst: NiMo on Alumina

Catalyst Life: 2 Years

## REACTOR OPERATING CONDITIONS:

	CASE	
	PHASE II MAX. JP-4	PHASE II MAX. JP-8
Feedstock TBP Boiling		
Range, °F	180-490	180-550
Total Nitrogen, ppm	3260	3480
LHSV, V/Hr/V	2.0	2.0
Avg. Catalyst Temp., °F	750	750
Total Pressure, psia	1500	1500
H <sub>2</sub> PP	1400	1400
Recycle Gas Rate, SCF/B	4000	4000
Hydrogen Consumption, SCF/B		
Chemical	350	400
Dissolved	50	50
Total to Hydrotreater	400	450
Product		
Total Nitrogen, ppm	8	8
Sulfur, ppm	2	2
C <sub>4</sub> + Yield, Vol.% Feed	101.66	100.97

TABLE 15

MATERIAL BALANCE SUMMARY FOR  
NAPHTHA HYDROTREATER (5000 ppm N<sub>T</sub>)

---

JP-4 OPERATION

Basis: 24,141 BPSD of 180-490°F Naptha Fraction

9.66 x 10<sup>6</sup> SCF Hydrogen PSD (8.45 x 10<sup>6</sup> SCH H<sub>2</sub> Chemically Consumed)

PRODUCTS

Ammonia, STSD	13.7	<u>INSPECTIONS ON 180-490°F. CUT</u>		
Hydrogen Sulfide, Sulfur, Eq, STSD	0.2		<u>FEED</u>	<u>PRODUCT</u>
Unreacted H <sub>2</sub> x 10 <sup>6</sup> SCF PSD	1.21	API Gravity @ 60°F	41.5	42.3
C <sub>1</sub> -C <sub>3</sub> Gases, Lbs PSD	13,956	Aromatics, Vol. %	24.3	15.0
		Olefins, Vol. %	3.0	1.4
C <sub>4</sub> -180°F, BPSD	1,207	Total Nitrogen, ppm	3260	8.0
180-490°F, BPSD	<u>23,335</u>	Sulfur, ppm	65	2.0
TOTAL LIQUIDS, BPSD	24,542			

TABLE 16

MATERIAL BALANCE SUMMARY FOR  
NAPHTHA HYDROTREATER (5000 ppm N<sub>T</sub>)

JP-8 OPERATION

BASIS: 27,995 BPSD of 180-550°F Kerosine Fraction

12.60 x 10<sup>6</sup> SCF Hydrogen PSD (11.20 x 10<sup>6</sup> SCF H<sub>2</sub> Chemically Consumed)

PRODUCTS

INSPECTIONS ON FEED AND PRODUCT

			<u>180-550°F FEED</u>	<u>290-550°F PRODUCT</u>
Ammonia, STSD	17.0			
Hydrogen Sulfide, Sulfur Eq, STSD	0.3	API Gravity @ 60°F	40.6	41.6
		Aromatics, Vol.%	25.0	15.0
Unreacted H <sub>2</sub> x 10 <sup>6</sup> SCF PSD	1.40	Olefins, Vol.%	3.2	1.4
C <sub>1</sub> -C <sub>3</sub> Gases, Lbs PSD	16,185	Total Nitrogen, ppm	3480	8.0
		Sulfur, ppm	80	2.0
C <sub>4</sub> -290°F, BPSD	4,937			
290-550°F, BPSD	<u>23,685</u>			
TOTAL LIQUIDS, BPSD	28,622			

TABLE 17  
OPERATING CONDITIONS FOR  
NAPHTHA HYDROTREATER (6400 ppm N<sub>T</sub>)

---

Operator Factor: 0.90

Catalyst: NiMo on Alumina  
Catalyst Life: 1.5 Years

REACTOR OPERATING CONDITIONS:

	CASE	
	<u>PHASE II MAX. JP-4</u>	<u>PHASE II MAX. JP-8</u>
Feedstock TBP Boiling		
Range, °F	180-490	180-550
Total Nitrogen, ppm	3940	4200
LHSV, V/Hr/V	2.0	2.0
Avg. Catalyst Temp., °F	760	760
Total Pressure, psia	1500	1500
H <sub>2</sub> PP	1400	1400
Recycle Gas Rate, SCF/B	4000	4000
Hydrogen Consumption, SCF/B		
Chemical	375	425
Dissolved	<u>50</u>	<u>50</u>
Total to Hydrotreater	425	475
Product		
Total Nitrogen, ppm	8	8
Sulfur, ppm	2	2
C <sub>4</sub> + Yield, Vol.% Feed	101.71	102.28

TABLE 18

MATERIAL BALANCE SUMMARY FOR  
NAPHTHA HYDROTREATER (6400 ppm N<sub>T</sub>)

---

JP-4 OPERATION

Basis: 19,808 BPSD of 180-490°F Naphtha Fraction

8.42 x 10<sup>6</sup> SCF Hydrogen PSD (7.43 x 10<sup>6</sup> SCH H<sub>2</sub> Chemically Consumed)

PRODUCTS

Ammonia, STSD	13.7	<u>INSPECTIONS ON 180-490°F. CUT</u>		
Hydrogen Sulfide, Sulfur, Eq, STSD	0.2		<u>FEED</u>	<u>PRODUCT</u>
Unreacted H <sub>2</sub> x 10 <sup>6</sup> SCF PSD	0.94	API Gravity @ 60°F	40.3	41.3
C <sub>1</sub> -C <sub>3</sub> Gases, Lbs PSD	11,451	Total Nitrogen, ppm	3940	8.0
C <sub>4</sub> -180°F, BPSD	990	Sulfur, ppm	85	2.0
290-550°F, BPSD	<u>19,157</u>			
TOTAL LIQUIDS, BPSD	20,147			



TABLE 19  
MATERIAL BALANCE SUMMARY FOR  
NAPHTHA HYDROTREATER (6400 ppm N<sub>T</sub>)

---

JP-8 OPERATION

Basis: 27,411 BPSD of 180-490°F Naphtha Fraction

13.02 x 10<sup>6</sup> SCF Hydrogen PSD (11.65 x 10<sup>6</sup> SCH H<sub>2</sub> Chemically Consumed)

PRODUCTS

Ammonia, STSD	20.4	<u>INSPECTIONS ON 180-490°F. CUT</u>		
Hydrogen Sulfide, Sulfur, Eq, STSD	0.4		<u>FEED</u>	<u>PRODUCT</u>
Unreacted H <sub>2</sub> x 10 <sup>6</sup> SCF PSD	1.37	API Gravity @ 60°F	38.6	39.8
C <sub>1</sub> -C <sub>3</sub> Gases, Lbs PSD	15,847	Total Nitrogen, ppm	4200	8.0
C <sub>4</sub> -180°F, BPSD	4,834	Sulfur, ppm	95	2.0
290-550°F, BPSD	<u>23,202</u>			
TOTAL LIQUIDS, BPSD	28,036			

TABLE 20

HCl EXTRACTION FOR REMOVING NITROGEN FROM  
HYDROTREATED SHALE OIL

JP-4 OPERATION

CASE

Total nitrogen in liquid effluent, ppm	<u>2200</u>	<u>5000</u>	<u>6400</u>
Hydrotreated Feed			
TBP Boiling Range, °F	490-1000	490-1000	490-1000
API Gravity	29.6	28.9	28.
Total Nitrogen, ppm	2400	4800	6887
Sulfur, ppm	107	140	506
Aromatics and Polars, Wt. %	---	42	---
Reactor Conditions (HCl Treatment)			
Residence Time, Minutes	30	30	30
Inlet Temp., °F	100	100	100
Outlet Temp., °F	110	110	110
Total Pressure, psig	1	1	1
HCL Addition, Lbs/100 lbs. Feed	1.10	2.68	2.25
Settling Time, Minutes	30	30	30
Raffinate Phase Data			
Yield, Wt. % Oil Charged	95.7	86.2	86.8
API Gravity	30.9	30.7	30.0
Total Nitrogen, ppm	650	700	1950
Sulfur, ppm	16	17	47
Aromatics and Polars, Wt. %	---	34	---
Chloride, ppm	174	406	204
HCl Adduct Decomposition Conditions			
Residence Time, Minutes	30	30	30
Temperature, °F	575	575	575
Total Pressure, psig	1	1	1
HCL Recovery, Wt. %	97.1	96.9	97.1
Decomposed Adduct (HCl-Free Basis)			
Yield, Wt. % Oil Charged	4.3	13.4	13.4
API Gravity	25.0	16.8	16.8
Total Nitrogen, Wt. %	4.10	13.4	13.4
Sulfur, ppm	2117	13.4	13.4
Aromatics and Polars, Wt. %	---	---	---
Chlorine, ppm	1000	---	---

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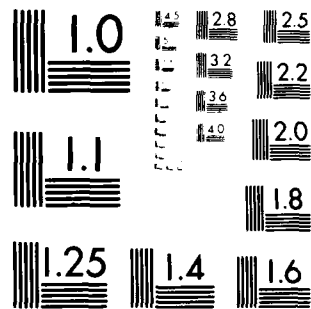
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TABLE 21  
MATERIAL BALANCE SUMMARY OF ANHYDROUS HYDROGEN CHLORIDE  
EXTRACTION (2200 ppm N<sub>T</sub>)

JP-4 OPERATION

Basis: 490-1000°F Hydrotreated Gas Oil Feed

	<u>Feed</u>	<u>Recovered HCl Free Raffinate</u>	<u>Recovered Extract</u>
Yields			
Wt. %	100	95.7	4.3
Vol. %	100	96.4	4.2
BPSD	71,904	69,281	3026
Inspections & Analyses			
API/Sp Grav. @ 60°F	29.6/0.8783	30.9/0.8714	25.0/0.9040
Total Nitrogen, ppm	2400	650	4.10 Wt. %
Sulfur, ppm	107	16	2117
Chlorine, ppm	0	174	200

Losses: Raffinate = 69 BPSD

Anhydrous HCL = 2363 Lbs/SD

TABLE 22

MATERIAL BALANCE SUMMARY OF ANHYDROUS HYDROGEN CHLORIDE  
EXTRACTION (5000 ppm N<sub>T</sub>)

---

JP-4 OPERATION

Basis: 490-1000°F Hydrotreated Gas Oil Feed

	<u>Feed</u>	<u>Recovered HCl-Free Raffinate</u>	<u>Recovered Extract</u>
<b>Yields</b>			
Wt. %	100	86.2	13.9
Vol. %	100	87.1	12.8
BPSD	73,133	63,681	9370
<b>Inspections &amp; Analyses</b>			
API/Sp Grav. @ 60°F	28.9/0.8823	30.7/0.8725	18.4/0.9542
Aromatics, wt. %	42	34	89
Total Nitrogen, ppm	4800	700	3.03 Wt. %
Sulfur, ppm	140	17	905
Chlorine, ppm	0	406	1500

Losses: Raffinate = 64 BPSD

Anhydrous HCl = 5872 Lbs/SD

**TABLE 23**  
**MATERIAL BALANCE SUMMARY OF ANHYDROUS HYDROGEN CHLORIDE**  
**EXTRACTION (6400 ppm N<sub>T</sub>)**

---

**JP-4 OPERATION**

**Basis: 490-1000°F Hydrotreated Gas Oil Feed**

	<u>Feed</u>	<u>Recovered HCl-Free Raffinate</u>	<u>Recovered Extract</u>
<b>Yields</b>			
Wt. %	100	88.8	11.2
Vol. %	100	89.8	10.1
BPSD	76,031	68,272	7703
<b>Inspections &amp; Analyses</b>			
API/Sp Grav. @ 60°F	28.1/0.8867	30.0/0.8760	12.4/0.9831
Total Nitrogen, ppm	6887	1950	4.59 Wt.%
Sulfur, ppm	306	47	2353
Chlorine, ppm	0	204	1200

**Losses: Raffinate = 68 BPSD**

**Anhydrous HCL = 5188 Lbs/SD**

TABLE 24

HCl EXTRACTION FOR REMOVING NITROGEN FROM  
HYDROTREATED SHALE OIL (5000 ppm)

JP-8 OPERATION

CASE

Total nitrogen in liquid effluent, ppm	<u>2200</u>	<u>5000</u>	<u>6400</u>
<b>Hydrotreated Feed</b>			
TBP Boiling Range, °F	550-1000	550-1000	550-1000
API Gravity	29.4	28.5	27.4
Total Nitrogen, ppm	2422	5600	7104
Sulfur, ppm	108	150	326
Aromatics and Polars, Wt.%	---	45	---
<b>Reactor Conditions (HCl Treatment)</b>			
Residence Time, Minutes	30	30	30
Inlet Temp., °F	100	100	100
Outlet Temp., °F	110	110	110
Total Pressure, psig	1	1	1
HCl Addition, Lbs/100 lbs. Feed	1.10	2.70	2.27
Settling Time, Minutes	30	30	30
<b>Raffinate Phase Data</b>			
Yield, Wt.% Oil Charged	95.7	86.2	88.9
API Gravity	30.6	30.3	29.7
Total Nitrogen, ppm	700	750	2000
Sulfur, ppm	75	28	27
Aromatics and Polars, Wt.%	---	35	---
Chloride, ppm	174	406	204
<b>HCl Adduct Decomposition Conditions</b>			
Residence Time, Minutes	30	30	30
Temperature, °F	575	575	575
Total Pressure, psig	1	1	1
HCl Recovery, Wt.%	97.1	96.9	97.4
<b>Decomposed Adduct (HCl-Free Basis)</b>			
Yield, Wt.% Oil Charged	4.3	13.9	11.1
API Gravity	24.8	16.3	10.5
Total Nitrogen, Wt.%	4.05	3.57	4.78
Sulfur, ppm	1897	928	2329
Aromatics and Polars, Wt.%	---	89	---
Chlorine, ppm	1000	1500	1500



TABLE 25  
MATERIAL BALANCE SUMMARY OF ANHYDROUS HYDROGEN CHLORIDE  
EXTRACTION (2200 ppm N<sub>T</sub>)

---

JP-8 OPERATION

Basis: 550-1000°F Hydrotreated Gas Oil Feed

	<u>Feed</u>	<u>Recovered HCl-Free Raffinate</u>	<u>Recovered Extract</u>
Yields			
Wt. %	100	95.7	4.3
Vol. %	100	96.3	4.2
BPSD	69,944	67,346	2943
Inspections & Analyses			
API/Sp Grav. @ 60°F	29.4/0.8793	30.6/0.8730	24.8/0.9052
Total Nitrogen, ppm	2422	700	4.05 Wt. %
Sulfur, ppm	108	27	1897
Chlorine, ppm	0	174	1000

Losses: Raffinate = 67 BPSD

Anhydrous HCl = 2302 Lbs/SD

TABLE 26

MATERIAL BALANCE SUMMARY OF ANHYDROUS  
HYDROGEN CHLORIDE EXTRACTION (5000 ppm N<sub>T</sub>)

JP-8 OPERATION

Basis: 550-1000°F Hydrotreated Gas Oil Feed

	<u>Feed</u>	<u>Recovered HCl-Free Raffinate</u>	<u>Recovered Extract</u>
<b>Yields</b>			
Wt. %	100	86.2	13.9
Vol. %	100	87.1	12.8
BPSD	69,279	60,329	8867
<b>Inspections &amp; Analyses</b>			
API/Sp Grav. @ 60°F	28.5/0.8842	30.3/0.8744	16.3/0.9573
Aromatics, Wt. %	45	35	89
Total Nitrogen, ppm	5600	750	3.57 Wt. %
Sulfur, ppm	150	28	908
Chlorine, ppm	0	406	1500

Losses: Raffinate = 60 BPSD

Anhydrous HCl = 5618 Lbs/SD

TABLE 27  
MATERIAL BALANCE SUMMARY OF ANHYDROUS  
HYDROGEN CHLORIDE EXTRACTION (6400 ppm N<sub>T</sub>)

JP-8 OPERATION

Basis: 550-1000°F Hydrotreated Gas Oil Feed

	<u>Feed</u>	<u>Recovered HCl-Free Raffinate</u>	<u>Recovered Extract</u>
<b>Yields</b>			
Wt. %	100	88.9	11.1
Vol. %	100	90.1	10.0
BPSD	68,428	61,637	6880
<b>Inspections &amp; Analyses</b>			
API/Sp Grav. @ 60°F	27.4/0.8904	29.7/0.8779	10.5/0.9862
Total Nitrogen, ppm	7104	2000	4.78 Wt.%
Sulfur, ppm	326	75	2329
Chlorine, ppm	0	204	1500

Losses: Raffinate = 62 BPSD

Anhydrous HCl = 4724 Lbs/SD

TABLE 28  
MAXIMUM JP-4 OPERATING CONDITIONS  
FOR GAS OIL HYDROCRACKER (2200 ppm N<sub>T</sub>)

<u>BASIS</u>	<u>PHASE II</u>	
	R-1	R-2
Reactor		
Operating Factor	0.90	
Catalyst	NiMo	"B"
Catalyst Life, Years	0.5	1.25
<u>REACTOR OPERATING CONDITIONS:</u>		
LHSV, vol/hr/vol	1.0	1.2
Average Catalyst Temp., °F	724	741
Total Pressure, psig	1700	
Recycle Gas Rate, SCF/B	6000	
Hydrogen Consumption, SCF/B		
Chemical	1251	
Dissolved	102	
Total	1353	
<u>FEEDSTOCK CHARACTERIZATION:</u>		
TBP Boiling Range, °F	490-1000	
API Gravity	30.9	
Total Nitrogen, ppm	650	
<u>PRODUCTS, VOL.% FRESH FEED:</u>		
C <sub>4</sub> + Yield	120.8	
JP-4	115.8	

TABLE 29

MATERIAL BALANCE SUMMARY FOR GAS OIL HYDROCRACKER  
JP-4 OPERATION (2200 ppm N<sub>T</sub>)

---

BASIS: 69,281 BPSD of HCl Raffinate

93.74 x 10<sup>6</sup> SCF Hydrogen PSD (86.67 x 10<sup>6</sup> SCF H<sub>2</sub> PSD  
Chemically Consumed)

PRODUCTS

Ammonium Chloride, STSD	2.8
Ammonia, STSD	7.4
Hydrogen Sulfide, Sulfur Eq. STSD	0.2
Unreacted H <sub>2</sub> x 10 <sup>6</sup> SCF PSD	7.07
C <sub>1</sub> -C <sub>3</sub> Gases, lbs. PSD	449,634
C <sub>4</sub> -180°F, BPSD	24,110
180-490°F, BPSD	56,118
490°F+ Recycle Drag Stream, BPSD	3,464

INSPECTIONS ON FEED AND PRODUCT

	<u>FEED</u>	<u>JP-4 PRODUCT</u>
API Gravity @ 60°F	30.9	53.0
Aromatics, %	-	13 vol.
Olefins, vol. %	-	1.4
Total Nitrogen, ppm	650	1
Sulfur, ppm	16	1

TABLE 30  
MAXIMUM JP-4 OPERATING CONDITIONS  
FOR GAS OIL HYDROCRACKER (5000 ppm N<sub>T</sub>)

BASIS	PHASE II	
	R-1	R-2
Reactor		
Operating Factor	0.90	
Catalyst	NiMo	"B"
Catalyst Life, Years	0.5	1.25
<u>REACTOR OPERATING CONDITIONS:</u>		
LHSV, vol/hr/vol	1.0	1.2
Average Catalyst Temp., °F	724	741
Total Pressure, psig	1700	
Recycle Gas Rate, SCF/B	6000	
Hydrogen Consumption, SCF/B		
Chemical	1258	
Dissolved	103	
Total	1361	
<u>FEEDSTOCK CHARACTERIZATION:</u>		
TBP Boiling Range, °F	490-1000	
API Gravity	30.7	
Total Nitrogen, ppm	700	
<u>PRODUCTS, VOL.% FRESH FEED:</u>		
C <sub>4</sub> + Yield	121.2	
JP-4	115.9	

TABLE 31

MATERIAL BALANCE SUMMARY FOR GAS OIL HYDROCRACKER  
JP-4 OPERATION (5000 ppm N<sub>T</sub>)

---

BASIS: 63,681 BPSD of HCl Raffinate

86.67 x 10<sup>6</sup> SCF Hydrogen PSD (80.11 x 10<sup>6</sup> SCF H<sub>2</sub> PSD  
Chemically Consumed)

PRODUCTS

Ammonium Chloride, STSD	6.0
Ammonia, STSD	6.3
Hydrogen Sulfide, Sulfur Eq. STSD	0.2
Unreacted H <sub>2</sub> x 10 <sup>6</sup> SCF PSD	6.56
C <sub>1</sub> -C <sub>3</sub> Gases, lbs. PSD	421,759
C <sub>4</sub> -180°F, BPSD	22,862
180-490°F, BPSD	50,881
490°F+ Recycle Drag Stream, BPSD	3,414

INSPECTIONS ON FEED AND PRODUCT

	<u>FEED</u>	<u>JP-4 PRODUCT</u>
API Gravity @ 60°F	30.7	52.6
Aromatics, %	34 wt.	14 vol.
Olefins, vol. %	-	1.4
Total Nitrogen, ppm	700	1
Sulfur, ppm	17	1

TABLE 32

MAXIMUM JP-4 OPERATING CONDITIONS  
FOR GAS OIL HYDROCRACKER (6400 ppm N<sub>T</sub>)

<u>BASIS</u>	<u>PHASE II</u>	
Reactor	R-1	R-2
Operating Factor	0.90	
Catalyst	NiMo	"B"
Catalyst Life, Years	0.5	1.25

REACTOR OPERATING CONDITIONS:

LHSV, vol/hr/vol	0.6	1.2
Average Catalyst Temp., °F	745	743
Total Pressure, psig	1700	
Recycle Gas Rate, SCF/B	6000	
Hydrogen Consumption, SCF/B		
Chemical	1381	
Dissolved	105	
Total	1486	

FEEDSTOCK CHARACTERIZATION:

TBP Boiling Range, °F	490-1000
API Gravity	30.0
Total Nitrogen, ppm	1950

PRODUCTS, VOL.% FRESH FEED:

C <sub>4</sub> + Yield	121.7
JP-4	116.7



TABLE 33

MATERIAL BALANCE SUMMARY FOR GAS OIL HYDROCRACKER  
JP-4 OPERATION (6400 ppm N<sub>T</sub>)

---

BASIS: 68,272 BPSD of HCl Raffinate

101.45 x 10<sup>6</sup> SCF Hydrogen PSD (94.28 x 10<sup>6</sup> SCF H<sub>2</sub> PSD  
Chemically Consumed)

PRODUCTS

Ammonium Chloride, STSD	3.2
Ammonia, STSD	23.7
Hydrogen Sulfide, Sulfur Eq. STSD	0.5
Unreacted H <sub>2</sub> x 10 <sup>6</sup> SCF PSD	7.17
C <sub>1</sub> -C <sub>3</sub> Gases, lbs. PSD	469,438
C <sub>4</sub> -180°F, BPSD	27,240
180-490°F, BPSD	52,433
490°F+ Recycle Drag Stream, BPSD	3,414

INSPECTIONS ON FEED AND PRODUCT

	<u>FEED</u>	<u>JP-4 PRODUCT</u>
API Gravity @ 60°F	30.0	51.8
Aromatics, vol.%		14
Olefins, vol.%	-	1.4
Total Nitrogen, ppm	1950	1
Sulfur, ppm	47	1

TABLE 34  
MAXIMUM JP-8 OPERATING CONDITIONS  
FOR GAS OIL HYDROCRACKER (2200 ppm N<sub>T</sub>)

<u>BASIS</u>	<u>PHASE II</u>	
Reactor	R-1	R-2
Operating Factor	0.90	
Catalyst	NiMo	"B"
Catalyst Life, Years	0.5	1.8

REACTOR OPERATING CONDITIONS:

LHSV, vol/hr/vol	1.0	1.2
Average Catalyst Temp., °F	724	709
Total Pressure, psig	1700	
Recycle Gas Rate, SCF/B	6000	
Hydrogen Consumption, SCF/B		
Chemical	962	
Dissolved	84	
Total	1046	

FEEDSTOCK CHARACTERIZATION:

TBP Boiling Range, °F	550-1000
API Gravity	30.6
Total Nitrogen, ppm	700

PRODUCTS, VOL.% FRESH FEED:

C <sub>4</sub> + Yield	117.1
C <sub>4</sub> -290°F	56.1
JP-8 (290-550°F B.R.)	56.0
550°F+ Bottoms	5.0

TABLE 35

MATERIAL BALANCE SUMMARY FOR GAS OIL HYDROCRACKER  
JP-8 OPERATION (2200 ppm N<sub>T</sub>)

---

BASIS: 67,346 BPSD of HCl Raffinate

70.44 x 10<sup>6</sup> SCF Hydrogen PSD (67.79 x 10<sup>6</sup> SCF H<sub>2</sub> PSD  
Chemically Consumed)

PRODUCTS

Ammonium Chloride, STSD	2.7
Ammonia, STSD	7.8
Hydrogen Sulfide, Sulfur Eq. STSD	0.3
Unreacted H <sub>2</sub> x 10 <sup>6</sup> SCF PSD	5.65
C <sub>1</sub> -C <sub>3</sub> Gases, lbs. PSD	114,488
C <sub>4</sub> -290°F, BPSD	37,781
290-550°F, BPSD	37,714
550°F + Recycle Drag Stream, BPSD	3,367

INSPECTIONS ON FEED AND PRODUCT

	<u>FEED</u>	<u>JP-8 PRODUCT</u>
API Gravity @ 60°F	30.6	40.2
Aromatics, vol.%		20
Olefins, vol.%	-	1.6
Total Nitrogen, ppm	700	1
Sulfur, ppm	27	1

TABLE 36  
MAXIMUM JP-8 OPERATING CONDITIONS  
FOR GAS OIL HYDROCRACKER (5000 ppm N<sub>T</sub>)

<u>BASIS</u>	<u>PHASE II</u>	
Reactor	R-1	R-2
Operating Factor	0.90	
Catalyst	NiMo	"B"
Catalyst Life, Years	0.5	1.8

REACTOR OPERATING CONDITIONS:

LHSV, vol/hr/vol	1.0	1.2
Average Catalyst Temp., °F	725	712
Total Pressure, psig	1700	
Recycle Gas Rate, SCF/B	6000	
Hydrogen Consumption, SCF/B		
Chemical	990	
Dissolved	85	
Total	1075	

FEEDSTOCK CHARACTERIZATION:

TBP Boiling Range, °F	550-1000
API Gravity	30.3
Total Nitrogen, ppm	750

PRODUCTS, VOL.% FRESH FEED:

C <sub>4</sub> + Yield	117.4
C <sub>4</sub> -290°F	56.9
JP-8 (290-550°F B.R.)	55.5
550°F+ Bottoms	5.0

TABLE 37

**MATERIAL BALANCE SUMMARY FOR GAS OIL HYDROCRACKER  
JP-8 OPERATION (5000 ppm N<sub>T</sub>)**

---

**BASIS: 60,329 BPSD of HCl Raffinate**

**64.85 x 10<sup>6</sup> SCF Hydrogen PSD (59.73 x 10<sup>6</sup> SCF H<sub>2</sub> PSD  
Chemically Consumed)**

**PRODUCTS**

Ammonium Chloride, STSD	5.7
Ammonia, STSD	6.6
Hydrogen Sulfide, Sulfur Eq. STSD	0.3
Unreacted H <sub>2</sub> x 10 <sup>6</sup> SCF PSD	5.12
C <sub>1</sub> -C <sub>3</sub> Gases, lbs. PSD	106,782
C <sub>4</sub> -290°F, BPSD	34,327
290-550°F, BPSD	33,483
550°F + Recycle Drag Stream, BPSD	3,016

**INSPECTIONS ON FEED AND PRODUCT**

	<u><b>FEED</b></u>	<u><b>JP-8 PRODUCT</b></u>
API Gravity @ 60°F	30.3	39.7
Aromatics, %	35 wt.	16 vol.
Olefins, vol. %	-	1.6
Total Nitrogen, ppm	750	1
Sulfur, ppm	28	1

TABLE 38  
MAXIMUM JP-8 OPERATING CONDITIONS  
FOR GAS OIL HYDROCRACKER (6400 ppm N<sub>T</sub>)

<u>BASIS</u>	<u>PHASE II</u>	
Reactor	R-1	R-2
Operating Factor	0.90	
Catalyst	NiMo	"B"
Catalyst Life, Years	0.5	1.8
<u>REACTOR OPERATING CONDITIONS:</u>		
LHSV, vol/hr/vol	0.6	1.2
Average Catalyst Temp., °F	746	709
Total Pressure, psig	1700	
Recycle Gas Rate, SCF/B	6000	
Hydrogen Consumption, SCF/B		
Chemical	1180	
Dissolved	90	
Total	1270	
<u>FEEDSTOCK CHARACTERIZATION:</u>		
TBP Boiling Range, °F	550-1000	
API Gravity	29.7	
Total Nitrogen, ppm	2000	
<u>PRODUCTS, VOL.% FRESH FEED:</u>		
C <sub>4</sub> + Yield	118.7	
C <sub>4</sub> -290°F	63.0	
JP-8 (290-550°F B.R.)	50.7	
550°F+ Bottoms	5.0	

TABLE 39  
MATERIAL BALANCE SUMMARY FOR GAS OIL HYDROCRACKER  
JP-8 OPERATION (6400 ppm N<sub>T</sub>)

---

BASIS: 61,637 BPSD of HCl Raffinate

78.28 x 10<sup>6</sup> SCF Hydrogen PSD (72.73 x 10<sup>6</sup> SCF H<sub>2</sub> PSD  
Chemically Consumed)

PRODUCTS

Ammonium Chloride, STSD	2.9
Ammonia, STSD	22.0
Hydrogen Sulfide, Sulfur Eq. STSD	0.7
Unreacted H <sub>2</sub> x 10 <sup>6</sup> SCF PSD	5.55
C <sub>1</sub> -C <sub>3</sub> Gases, lbs. PSD	141,765
C <sub>4</sub> -290°F, BPSD	38,831
290-550°F, BPSD	31,250
550°F + Recycle Drag Stream, BPSD	3,082

INSPECTIONS ON FEED AND PRODUCT

	<u>FEED</u>	<u>JP-8 PRODUCT</u>
API Gravity @ 60°F	29.7	39.8
Aromatics, vol.%	-	21
Olefins, vol.%	-	1.6
Total Nitrogen, ppm	2000	1
Sulfur, ppm	75	1

TABLE 40

**JP-4 AND OTHER FUELS - OPERATING CONDITIONS FOR  
GAS OIL HYDROCRACKER (2200 ppm N<sub>T</sub>)**

<u>BASIS:</u>	<u>PHASE II</u>	
Reactor	R-1	R-2
Operating Factor	0.90	
Catalyst	NiMo	"B"
Catalyst Life, Years	0.5	1.8

REACTOR OPERATING CONDITIONS:

LHSV, vol/hr/v	1.0	1.0
Average Catalyst Temp., °F	724	712
Total Pressure, psig	1700	
Recycle Gas Rate, SCF/B	6000	
Hydrogen Consumption, SCF/B		
Chemical	809	
Dissolved	78	
Total	887	

FEEDSTOCK CHARACTERIZATION

TBP Boiling Range, °F	490-1000
API Gravity	30.9
Total Nitrogen, ppm	650

PRODUCTS, VOL.% FRESH FEED

C <sub>4</sub> + Yield	115.1
JP-4	80.3
DF-2 (490-675°F B.R.)	29.6
Fuel Oil (675°F+ Bottoms)	5.2



TABLE 41

MATERIAL BALANCE SUMMARY FOR GAS OIL HYDROCRACKER  
JP-4 PLUS OTHER FUELS (2200 ppm N<sub>T</sub>)

---

BASIS: 69,281 BPSD of HCl Raffinate

61.45 x 10<sup>6</sup> SCF Hydrogen PSD (56.05 x 10<sup>6</sup> SCF H<sub>2</sub> PSD  
Chemically Consumed)

PRODUCTS

Ammonium Chloride, STSD	2.7
Ammonia, STSD	7.4
Hydrogen Sulfide, Sulfur Eq. STSD	0.2
Unreacted H <sub>2</sub> x 10 <sup>6</sup> SCF PSD	5.40
C <sub>1</sub> -C <sub>3</sub> Gases, lbs. PSD	103,922
C <sub>4</sub> -180°F, BPSD	11,639
180-490°F, BPSD	43,993
490-675°F, BPSD	20,507
675°F+ Bottoms, BPSD	3,602

INSPECTIONS ON FEED AND PRODUCTS

	<u>FEED</u>	<u>JP-4</u>	<u>DF-2</u>	<u>675°F+ BOTTOMS</u>
API Gravity @ 60°F	30.9	51.9	36.2	26.3
Aromatics, %	--	13 vol.	13 wt.	42.8 vol.
Olefins, vol. %	--	1.4	--	--
Total Nitrogen, ppm	650	1	1	3
Sulfur, ppm	16	1	1	2

TABLE 42  
JP-4 AND OTHER FUELS - OPERATING CONDITIONS FOR  
GAS OIL HYDROCRACKER (5000 ppm N<sub>T</sub>)

<u>BASIS:</u>	<u>PHASE II</u>	
Reactor	R-1	R-2
Operating Factor	0.90	
Catalyst	NiMo	"B"
Catalyst Life, Years	0.5	1.8

REACTOR OPERATING CONDITIONS:

LHSV, vol/hr/v	1.0	1.0
Average Catalyst Temp., °F	725	712
Total Pressure, psig	1700	
Recycle Gas Rate, SCF/B	6000	
Hydrogen Consumption, SCF/B		
Chemical	843	
Dissolved	80	
Total	923	

FEEDSTOCK CHARACTERIZATION

TBP Boiling Range, °F	490-1000
API Gravity	30.7
Total Nitrogen, ppm	700

PRODUCTS, VOL.% FRESH FEED

C <sub>4</sub> + Yield	116.7
JP-4	80.3
DF-2 (490-675°F B.R.)	29.2
Fuel Oil (675°F+ Bottoms)	7.2

TABLE 43

**MATERIAL BALANCE SUMMARY FOR GAS OIL HYDROCRACKER  
JP-4 PLUS OTHER FUELS (5000 ppm N<sub>T</sub>)**

---

**BASIS: 63,681 BPSD of HCl Raffinate**

**58.78 x 10<sup>6</sup> SCF Hydrogen PSD (53.68 x 10<sup>6</sup> SCF H<sub>2</sub> PSD  
Chemically Consumed)**

**PRODUCTS**

Ammonium Chloride, STSD	6.0
Ammonia, STSD	6.3
Hydrogen Sulfide, Sulfur Eq. STSD	0.2
Unreacted H <sub>2</sub> x 10 <sup>6</sup> SCF PSD	5.10
C <sub>1</sub> -C <sub>3</sub> Gases, lbs. PSD	97,432
C <sub>4</sub> -180°F, BPSD	10,507
180-490°F, BPSD	40,628
490-675°F, BPSD	18,595
675°F+ Bottoms, BPSD	4,585

**INSPECTIONS ON FEED AND PRODUCTS**

	<u>FEED</u>	<u>JP-4</u>	<u>DF-2</u>	<u>675°F+ BOTTOMS</u>
API Gravity @ 60°F	30.7	51.1	35.8	26.4
Aromatics, %	34 wt.	14 vol.	13 wt.	42.5 vol
Olefins, vol.%	--	1.4	--	--
Total Nitrogen, ppm	700	1	1	3
Sulfur, ppm	17	1	1	2

TABLE 44

**JP-4 AND OTHER FUELS - OPERATING CONDITIONS FOR  
GAS OIL HYDROCRACKER (6400 ppm N<sub>T</sub>)**

<u>BASIS:</u>	<u>PHASE II</u>	
Reactor	R-1	R-2
Operating Factor	0.90	
Catalyst	NiMo	"B"
Catalyst Life, Years	0.5	1.8

REACTOR OPERATING CONDITIONS:

LHSV, vol/hr/v	0.6	1.0
Average Catalyst Temp., °F	743	709
Total Pressure, psig	1700	
Recycle Gas Rate, SCF/B	6000	
Hydrogen Consumption, SCF/B		
Chemical	999	
Dissolved	84	
Total	1083	

FEEDSTOCK CHARACTERIZATION

TBP Boiling Range, °F	490-1000
API Gravity	30.0
Total Nitrogen, ppm	1950

PRODUCTS, VOL.% FRESH FEED

C <sub>4</sub> + Yield	118.7
JP-4	88.6
DF-2 (490-675°F B.R.)	23.1
Fuel Oil (675°F+ Bottoms)	7.0

TABLE 45

MATERIAL BALANCE SUMMARY FOR GAS OIL HYDROCRACKER  
JP-4 PLUS OTHER FUELS (6400 ppm N<sub>T</sub>)

---

BASIS: 68,272 BPSD of HCl Raffinate

73.94 x 10<sup>6</sup> SCF Hydrogen PSD (68.20 x 10<sup>6</sup> SCF H<sub>2</sub> PSD  
Chemically Consumed)

PRODUCTS

Ammonium Chloride, STSD	3.2
Ammonia, STSD	23.7
Hydrogen Sulfide, Sulfur Eq. STSD	0.5
Unreacted H <sub>2</sub> x 10 <sup>6</sup> SCF PSD	5.74
C <sub>1</sub> -C <sub>3</sub> Gases, lbs. PSD	129,717
C <sub>4</sub> -180°F, BPSD	14,679
180-490°F, BPSD	45,811
490-675°F, BPSD	15,770
675°F+ Bottoms, BPSD	4,779

INSPECTIONS ON FEED AND PRODUCTS

	<u>FEED</u>	<u>JP-4</u>	<u>DF-2</u>	<u>675°F+ BOTTOMS</u>
API Gravity @ 60°F	30.0	51.3	36.7	26.9
Aromatics, %	--	14 vol.	12 vol.	44 vol.
Olefins, vol. %	--	1.4	--	--
Total Nitrogen, ppm	1950	1	1	3
Sulfur, ppm	47	1	1	2

TABLE 46

## FINAL PRODUCT INSPECTIONS AND ANALYSES

CHEMICAL AND PHYSICAL TEST DATA	JP-4	JP-8	I-290°F (1)	DF-2	BLENDED HEAVY FUEL
API @ 60°F	50.9	42.9	71.0	38.0	25.7
Distillation, °F	D-2887	D-2887	D-86	D-86	---
I	158	210	50	450	---
10 vol. %	---	310	145	480	---
20 vol. %	260	352	170	495	---
50 vol. %	342	410	202	535	---
90 vol. %	459	510	240	590	---
EP	527	560	290	650	---
Aromatics, %	15	16	3	24	36
Olefins, vol. %	1	2	1	---	---
Mercaptans, wt. %	0.0001	0.0003	---	---	---
Sulfur, wt. %	0.0003	0.0002	NA	0.0011	0.06
Nitrogen (Total), ppm	5	3	1	3	3600
Flash Pt., °F	---	100	---	210	---
Freeze Pt., °F	-72	-70	-76	-10	---
Net Ht. of Comb., BTU/lb	18,764	18,610	19,050	18,730	18,500
H <sub>2</sub> Content, wt. %	14.16	13.85	NA	NA	NA

(1) RON and MON on unleaded fuel were 69.5 and 67.9 respectively.

TABLE 47  
BASIS FOR PHASE IV ECONOMICS

CAPITAL INVESTMENT

Plant Location	-	Salt Lake City, Utah
Refinery Capacity	-	100,000 BPSD raw shale oil
Cost Base	-	1st Quarter 1981
Plant Offsites	-	45% plant onsites minus cost of specified tankage
Financing	-	100% equity
	-	Three-year plant construction period
		25% first year, 50% second year, 25% third year
		Investment tax credit @ 10%

WORKING CAPITAL

Crude Inventory	-	21 days storage capacity/14 day inventory
Product Inventory	-	14 days storage capacity/ 7 day inventory
Crude Shale Oil	-	\$40.00 per barrel
Product Price	-	Product valued at actual cost; inventory at \$50.00 per barrel
Debt Financing	-	15% (including cost of initial catalyst loading)

CAPITAL RETURN

Discounted Cash Flow Rate	-	15%
Plant Salvage Value	-	Zero
Plant Depreciation	-	13 years sum of years digits

TABLE 47 (Continued)  
BASIS FOR PHASE IV ECONOMICS

OPERATING BASES

Plant Life	- 16 years
Plant Operating Factors	- 50% operating capacity 1st year
Plant On Stream Factor	- 90% after 1st year

OPERATING COST BASES

Process Heat	- Requirements Generated Internally
Cooling Water	- 3¢/1000 Gallons
Electricity	- 4.5¢ KWHR
Operator <sup>(1)</sup>	- \$12.00/manhour
Helpers <sup>(1)</sup>	- \$10.50/manhour
Supervision	- 25% of direct labor
Overhead	- 100% of direct labor
Taxes	- federal & state combined @ 50%
Maintenance, Local Taxes & Insurance	- 4.5% of fixed investment
Product Values	- all fuels are equal value
By-Product Values	- ammonia \$155.00 per short ton Sulfur \$105.00 per long ton

(1) Based on 4.2 shift positions plus 10% relief for continuous operation.



TABLE 48  
MATERIAL BALANCE SUMMARY

BASIS: OPTIMIZED 100,000 BPSD RAW OCCIDENTAL SHALE OIL REFINERY

	<u>MAX. JP-4</u>	<u>MAX. JP-8</u>	<u>JP-4 PLUS OTHER FUELS</u>
Net Products, BPSD (TBP Cuts)			
C4-290°F B.R. Naphtha	---	43,716	---
C4-490°F B.R. JP-4	108,504	---	83,810
290-550°F B.R. JP-8	---	62,141	---
550°F+ Recycle Drag	---	3,367	---
490°F+ HC Recycle Drag	3,464	---	---
490-675°F B.R. DF-2	---	---	16,454
675-1000°F B.R. Heavy Fuel	---	---	5,777
TOTAL FUELS	111,968	109,224	106,141
Other Products, STPSD			
Liquid Ammonia	245.9	254.2	199.4
Sulfur	119.6	112.5	110.7
Ammonium Chloride	2.8	2.7	3.2
Liquid Fuel Yields			
Products, vol.% feed to Raw Shale Oil Hydrotreater			
Naphtha	---	43.7	---
JP-4	108.5	---	83.9
JP-8	---	62.1	---
DF-2	---	---	16.5
Heavy Fuel	3.5	3.4	5.8
Total Refinery Input (crude, fuel & utilities converted to FOE), BPSD	114,973	114,334	105,677
Products, vol. % Total Refinery Input			
Naphtha	---	38.2	---
JP-4	94.4	---	79.0
JP-8	---	55.2	---
DF-2	---	---	15.6
Heavy Fuel	3.0	2.9	5.5
Overall Refinery thermal Energy Efficiency, %	81.5	80.7	86.7

TABLE 49

TOTAL HYDROGEN CHEMICALLY CONSUMED IN OPTIMIZED CASES

BASIS: 100,000 BPSD Raw Occidental Shale Oil Feedstock to the  
Main Hydrotreater

	SCF 100% HYDROGEN X 10 <sup>6</sup> CONSUMED PSD		
	<u>MAX JP-4</u>	<u>MAX JP-8</u>	<u>JP-4 &amp; OTHER FUELS</u>
Raw Shale Oil Hydrotreater Effluent Severity, ppm N <sub>T</sub> in Liquid	2200	2200	6400
Raw Shale Hydrotreater	157.0	157.0	120.0
Naphtha Hydrotreater	7.6	8.8	8.4
Gas Oil Raffinate Hydrocracker	93.8	70.5	67.6
Total	258.4	236.3	196.0
Total Hydrogen Consumed, SCF per Bbl of Raw Shale Oil	2584	2363	1960

TABLE 50

PLANT CAPACITIES AND ESTIMATED FIRST QUARTER 1981 INVESTMENTS (PHASE IV)

	MAX. JP-4		MAX. JP-8		JP-4 + OTHER FUELS	
	CAPACITY PSD	\$ x 10 <sup>6</sup>	CAPACITY PSD	\$ x 10 <sup>6</sup>	CAPACITY PSD	\$ x 10 <sup>6</sup>
H <sub>2</sub> Plant (TPO), MMSCF/SD (100% H <sub>2</sub> basis)	155.9	120.8	174.5	127.5	139.4	112.4
H <sub>2</sub> Plant (steam reforming), MMSCF/SD (100% H <sub>2</sub> Basis)	102.5	39.5	61.8	29.0	56.7	27.9
Sulfur Recovery, STSD	112.2	11.3	112.5	11.0	110.7	10.9
Waste Water Treating, STSD,	246.0	15.7	246.4	16.0	119.4	14.0
NH <sub>3</sub>						
Main Hydrotreater & H <sub>2</sub> S Recovery, MBPSD	100.0	183.8	100.0	183.8	100.0	183.8
Atm. & Vac. Distn., MBPSD	103.9	49.7	103.9	49.7	103.9	49.7
Dist. Hydrotreater, MBPSD	27.1	37.3	29.1	47.8	19.8	30.9
HCl Treater, MBPSD	71.9	2.6	69.9	2.5	76.0	2.7
Hydrocracker & Atm. Distn., MBPSD Fresh Feed	69.3	<u>107.7</u>	67.3	<u>103.5</u>	68.3	<u>97.7</u>
Subtotal		581.4		570.5		530.0
Tankage, MM BBLs.	3.6	<u>35.6</u>	3.6	<u>35.3</u>	3.6	<u>35.6</u>
Total On-Sites		617.0		605.8		565.6
Off-Sites (45% on-sites minus tankage)		<u>261.6</u>		<u>256.7</u>		<u>238.5</u>
Total Capital Investment		878.6		862.5		804.1

TABLE 51

PHASE IV COST COMPARISON FOR MANUFACTURING MILITARY FUELS  
FROM RAW OCCIDENTAL SHALE OIL

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BASIS: OPTIMIZED 100,000 BPSD REFINERY CRUDE CAPACITY (90,000 BPCD)

<u>CASE</u>	<u>MAX. JP-4</u>	<u>MAX. JP-8</u>	<u>JP-4 AND OTHER FUELS</u>
<u>TOTAL PLANT INVESTMENT, \$ 10<sup>6</sup></u>			
Plant	878.6	862.5	804.1
Catalysts	19.3	19.0	16.8
Working Capital	<u>98.1</u>	<u>97.1</u>	<u>91.0</u>
TOTAL	996.0	978.6	911.9
<u>MANUFACTURING COSTS - \$/CD</u>			
Direct Labor	15,538	15,538	15,538
Purchased Power and Cooling Water	91,140	91,701	74,081
Catalyst, Chemicals & Royalties	47,052	47,160	44,038
Overhead @ 100% Direct Labor	15,538	15,538	15,538
Maint., Local Taxes & Insurance	76,064	74,689	69,730
Depreciation (Average 13 years)	188,535	185,121	172,835
Subtotal	433,867	429,747	391,760
Less NH <sub>3</sub> & S (Credit)	(50,673)	(51,214)	(42,531)
Direct Costs	383,194	378,533	349,229
Liquid Product, \$/Bbl	3.80	3.85	3.66
TOTAL LIQUID FUELS, BPCD	100,771	98,306	95,527
TOTAL MANUFACTURING COSTS, \$/Bbl Product <sup>1</sup>	10.30	10.38	9.94
Adjusted Crude Cost, \$/Bbl Product	<u>41.07</u>	<u>41.87</u>	<u>39.83</u>
TOTAL PRODUCT COST			
\$/Bbl	51.37	52.25	49.77
¢/Gal	122.3	124.4	118.5

<sup>1</sup> Total Manufacturing Costs Computed on the Basis Shown in Table 43  
for Developing Phase IV Economics

TABLE 52  
SUMMARY OF SUN TECH'S OPTIMIZED  
PROCESSING SCHEMES

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BASIS: PHASE IV ECONOMICS

Processing Route For	<u>MAX. JP-4</u>	<u>MAX. JP-8</u>	<u>JP-4 AND OTHER FUELS</u>
Raw Shale Oil Input to Main Hydrotreater, BPSD	100,000	100,000	100,000
Total Refinery Input <sup>1</sup>	114,973	114,334	105,677
Products, BPSD			
Jet Fuel	108,504	62,141	83,910
Total Liquid Products	111,968	109,224	106,141
Liquid Fuel Yields as Vol % Crude Processed			
Jet Fuel	108.5	62.1	83.9
Total Fuels	112.0	109.2	106.1
Products as vol % Total Energy Input			
Jet Fuel	94.4	54.3	79.1
Total Liquid Products	97.4	95.5	100.7
Product Cost, \$/B	51.37	52.25	49.77
¢/gal	122.3	124.4	118.5
Overall Thermal Efficiency, %	81.5	80.7	86.7
Plant Investment, \$/SDB	8786	8625	8041

<sup>1</sup> Crude + Fuel + Utilities converted to FOE basis.

TABLE 53  
SENSITIVITY OF FUEL COST TO CHANGES  
IN INTEREST RATE OF RETURN (IRR)

<u>IRR. %</u>	<u>ESTIMATED FUEL COST (CENTS/GAL)</u>		
	<u>MAX. JP-4</u>	<u>MAX. JP-8</u>	<u>JP-4 PLUS OTHER FUELS</u>
10	115.2	117.3	111.6
15 (Base)	122.3	124.4	118.7
20	131.4	133.6	127.2

TABLE 54  
SENSITIVITY OF FUEL COST TO CHANGES  
IN PRICE OF RAW SHALE OIL

<u>Raw Shale Oil Price, \$/Bbl</u>	<u>ESTIMATED FUEL COST (CENTS/GAL)</u>		
	<u>MAX. JP-4</u>	<u>MAX. JP-8</u>	<u>JP-4 PLUS OTHER FUELS</u>
35	110.0	111.8	106.5
40 (Base)	122.3	124.4	118.7
45	134.7	137.0	130.4
50	147.0	149.8	142.4

TABLE 55

SENSITIVITY OF FUEL COST TO  
CHANGES IN TOTAL PLANT INVESTMENT

<u>CAPITAL INVESTMENT %</u>	<u>ESTIMATED FUEL COST (CENTS/GAL)</u>		
	<u>MAX. JP-4</u>	<u>MAX. JP-8</u>	<u>JP-4 PLUS OTHER FUELS</u>
90	120.5	122.6	116.7
100 (Base)	122.3	124.4	118.7
110	124.1	126.2	120.3
125	126.8	129.0	123.1

TABLE 56

SENSITIVITY OF FUEL COST TO CHANGES  
IN ANNUAL INTEREST RATE FOR WORKING CAPITAL

<u>ANNUAL INTEREST RATE %</u>	<u>ESTIMATED FUEL COST (CENTS/GAL)</u>		
	<u>MAX. JP-4</u>	<u>MAX. JP-8</u>	<u>JP-4 PLUS OTHER FUELS</u>
10	121.9	124.0	118.2
15 (Base)	122.3	124.4	118.7
20	122.7	124.8	118.8

TABLE 57  
EFFECT OF INTEREST RATE ON BORROWED CAPITAL  
ON PRODUCT COSTS

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BASIS: PHASE IV ECONOMICS BASIS

<u>CASE</u>	<u>MAX. JP-4</u>	<u>MAX. JP-8</u>	<u>JP-4 AND OTHER FUELS</u>
Plant Investment, \$ x 10 <sup>6</sup>	878.6	862.5	804.1
Total Liquid Fuels, BPCD	111,968	109,224	106,141
<u>Base</u>			
Product Cost, ¢/Gallon Working Capital Only @ 15% Interest Rate (Base)	122.3	124.4	118.5
Working Capital and 100% Plant Investment @ 15% Interest Rate, ¢/Gallon	131.7	133.9	127.6
ΔCost, ¢/Gallon	9.4	9.5	9.1
@ 20% Interest Rate, ¢/Gallon	135.3	137.5	131.0
ΔCost, ¢/Gallon	13	13.1	12.5



## APPENDIX A

### FLUID CATALYTIC CRACKING RESULTS

#### SUMMARY

This appendix reports the results of a brief fluid catalytic cracking (FCC) study made on the 480°F+ bottoms material obtained from our JP-4 production run at Hydrocarbon Research, Inc.<sup>(9)</sup> The results indicate that this fraction from severely hydrotreated Geokinetics shale oil would make a suitable FCC feedstock without any additional treatment. However, the front end has a lower distillation range than conventional cat cracker feedstocks; a more realistic feedstock would be a 600°F+ bottoms fraction from the severely hydrotreated Geokinetics shale oil.

At 80% conversion, this feedstock gave C<sub>5</sub>+ gasoline and coke yields of about 60 vol.% and 1.9 wt.% of fresh feed, respectively. Clear research and motor octanes of 83.9 and 77.7 were lower than expected. These low octanes are probably related to the unusual feedstock characteristics--low initial boiling point (333°F), high n-paraffin and low aromatics contents and a 950°F average FCC reactor temperature.

#### EXPERIMENTAL DETAILS

Results of catalytic cracking hydrotreated Geokinetic shale oil (480°F+ bottoms) are summarized in Tables A-1 and A-2. The data were generated at pilot plant conditions chosen to predict approximate yields using an equilibrium catalyst obtained from a commercial unit.

The hydrotreated shale oil is readily cracked as the feedstock inspections and yield data show. Compared to gas oils obtained from a conventional crude oil, the hydrotreated shale oil has a lower basic nitrogen level, higher API gravity, lower Ramsbottom carbon and lower aromatics content. A comparison with "typical" feedstock data is shown below:

	<u>480°F+ Hydrotreated Shale Oil</u>	<u>"Typical" Cracker Feedstock</u>
Basic Nitrogen, ppm	66.8	200 - 300
API Gravity	38.1	24 - 30
Ramsbottom Carbon, Wt. %	0.06	0.2 - 0.5
Aromatics, Wt. %	20.5	30 - 40
Pour Point, °F	+80	+60

The 80 vol.% conversion is typical for the pilot plant conditions that were chosen. This conversion level had been previously achieved with a petroleum derived feedstock using the same catalyst and identical pilot plant severity. (See Table A-2)

The pilot plant data predict yields; however, the low coke make (~ 2 wt. % of fresh feed) is not practical in a commercial heat balanced unit unless thermal requirements are satisfied by techniques such as CO combustion (high temperature catalyst regeneration), feed preheat, or regenerator torch oil injection. Since the shale oil cracking was once through, more coke would have been made if a recycling operation were practiced. However, it is still doubtful that there would be enough coke produced with this light feed to satisfy the heat balance requirements.

The C<sub>5</sub>+ gasoline sample had F-1 and F-2 clear octane numbers of 83.9 and 77.7 respectively, which are lower than the octane numbers obtained from a petroleum derived feedstock. The low research and motor octane numbers of the C<sub>5</sub>+ product gasoline were probably affected by the low-boiling front end of the feedstock. This front end material is refractory, has a low octane value and overlaps the heavy end of the catalytic gasoline boiling range. Hence, on distillation from the cracked product it is included in the catalytic gasoline fraction. A catalytic cracking feedstock with a 600°F initial boiling point would have produced a higher octane gasoline. A feedstock with a 600°F IBP would comprise only 50 vol.% of the 480°F+ bottoms. There are options available that were not explored in this preliminary work that would increase the octane number in the catalytic gasoline.

This hydrotreated shale oil was very waxy, had less aromatics and had a lighter front end than conventional feedstocks from petroleum. Trends show that the more aromatic the feedstock, the higher the expected octane. One additional factor that may partially account for the lower octane number is the difference in operating pressures. The pressure was 7 psi higher than the normal operating pressure (20 psig).

TABLE A-1  
INSPECTIONS AND ANALYSES OF HYDROGENATED 480°F+  
BOTTOMS FROM GEOKINETICS SHALE OIL

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Gravity, °API @ 60°F	38.1
V.B.R., °F (Converted to 1 Atm.)	
IBP	333
5	485
10	496
20	523
30	550
40	578
50	606
60	642
70	688
80	740
90	801
95	838
EP	878
% Recovery	98
Sulfur, ppm	24
Total Nitrogen, ppm	109
Basic Nitrogen, ppm	67
Refractive Index @ 67°F	1.4451
Specific Gravity @ 60°F	0.8343
Average Molecular Weight	325
Aromatics, Wt. %	20.5
Ramsbottom Carbon, Wt. %	0.06
Vis. SUS @ 100°F	43.6
@ 210°F	32.2
Pour Pt., °F	+80

TABLE A-2  
YIELDS FROM CATALYTIC CRACKING 480°F+ BOTTOMS FROM  
HYDROTREATED GEOKINETICS SHALE OIL

Reactor Operating Conditions

Reactor Temp., °F	950
Cat/Oil Ratio, wt/wt	5.97
Oil Contact Time, sec.	6.35
Combined Feed Ratio, vol/vol	1.00
Pressure, psig	27
Material Balance, wt.% Feed	97.9

Yields, Vol.% Fresh Feed (Normalized to 100 Wt.% Feed)

	<u>Shale Bottoms</u>	<u>Petroleum Derived Feedstock</u>
H <sub>2</sub> (FOE)	.13	.14
C <sub>1</sub> "	.60	1.34
C <sub>2</sub> "	.78	1.07
C <sub>2</sub> "	.57	.92
C <sub>3</sub>	7.84	8.08
C <sub>3</sub>	3.59	3.71
C <sub>4</sub>	5.35	3.88
iC <sub>4</sub>	11.79	10.1
nC <sub>4</sub>	3.01	2.22
C <sub>5</sub> + Gasoline (90% @ 385°F)	60.41	59.45
F-1 Octane	83.9	90.6
F-2 Octane	77.7	80.2
Sensitivity	6.2	10.4
Bottoms	20.0	19.55
Total Liquids, vol.% Fresh Feed	114.08	111.12
Coke, Wt. % Fresh Feed	1.9	5.66
Conversion, Vol. % Fresh Feed <sup>(1)</sup>	80.00	80.45
CO <sub>2</sub> /CO in flue gas	5.42	4.11

(1) Conversion = 100 minus vol.% bottoms

## APPENDIX B

### PARAFFINIC BASE OILS FROM HYDROGENATED SHALE OIL

#### SUMMARY

As part of our evaluation of potential applications for the hydrogenated 480°F+ bottoms fraction from Geokinetics shale oil from the JP-4 production run at Hydrocarbon Research, Inc.<sup>(9)</sup> some preliminary screening tests were made to determine whether this material might be suitable for use in the manufacture of paraffinic base oils. These results indicate that this material may be acceptable for making 130 SUS base oils of about 100 VI and 0°F pour point at a yield of about 11 vol.% of the 480°F+ bottoms. Paraffin wax would be a co-product of the lube oil refining process. The quality of the wax was not determined.

#### EXPERIMENTAL DETAILS

About 850 barrels of Geokinetics shale oil was severely hydrogenated to produce specification JP-4 turbine fuel using Sun Tech's upgrading technology.<sup>(9)</sup> The bottoms fraction from this operation was quite waxy, and it was thought that it might be a suitable feedstock for manufacturing paraffinic base oils. Table B-1 gives the inspections obtained on a sample of solvent lube and slack wax prepared from the 720°F+ bottoms material from a vacuum distillation. Note that the end point is under 900°F, hence the maximum potential lube oil viscosity would be low.

The 480°F+ bottoms was distilled into an overhead and 720°F+ bottoms fractions. An abbreviated solvent refining examination was carried out on the 720°F+ bottoms fraction to estimate its lubricating oil potential

and quality. Solvent lube processing yield estimates indicate the following:

Lube Yield on 480°F+ Bottoms, vol.%	11.0
API Gravity	33.0
Viscosity, SUS @ 100°F	134
VI	112
Aromatics, wt.%	18
Pour Point, °F	0

Since this paraffinic base oil sample was prepared from a hydrocracked stock, an additional finishing step would be needed to make it stable to both oxidation and exposure to ultra violet light. The aromatics content of the base oil fraction is similar to that normally found in comparable solvent refined paraffinic base oils.

Additional development work would be needed to insure that quality base oils could be made in acceptable yields by this processing route from raw shale oils. The slack wax from the dewaxing step would require further development work to determine its value and quality for paraffin wax applications.

TABLE B-1

Preliminary Solvent Lube Screening Evaluation of  
Hydrotreated Geokinetics Shale Oil

Distillation, Vol. %

IBP-720°F Distillate	77.40
720°F + Waxy Bottoms	22.60

Estimated Solvent Lube Yield

Base Oil Yield, Vol. % 480°F+ Bottoms	11.0
Vol. % 720°F+ Bottoms	45.9
Viscosity, SUS @ 100°F (Centistokes)	134.3 (28.4)
@ 210°F (Centistokes)	42.7 (5.01)
VI	112
Aromatics, Wt. %	18
Pour Point, °F	0
Slack Wax Yield @ 20% ED Oil Content	
Vol. % 480°F+ Btms.	11.6
Vol. % 720°F+ Btms.	51.1



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